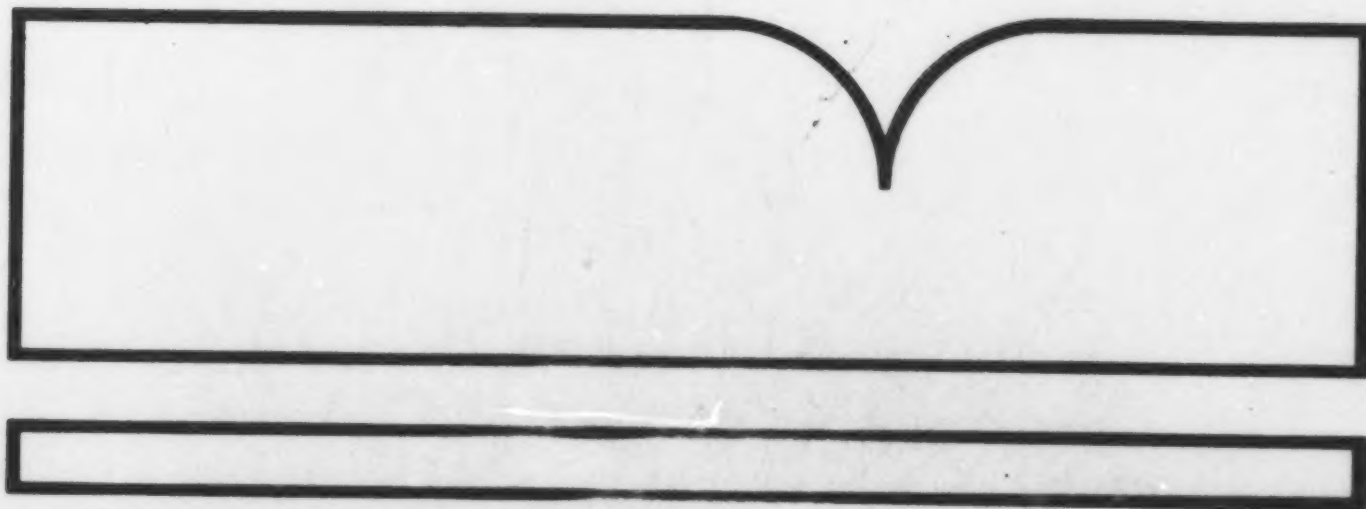


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Technical Report

Indirect Liquefaction Processes

by

John McGuckin

February 1982

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16. ABSTRACT This report examines the technology feasibility of the various coal gasification and indirect liquefaction technologies. Also included is the best-estimate costs for methanol and gasoline using the various technologies with three different coal/feedstocks by critically analysing publicly available design studies and placing them on a common technical/financial basis. The following conclusion is that methanol from coal is cheaper than gasoline via either the Mobile MTG process or the Fisher/Tropsch process.		
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I. Introduction/Summary

The purpose of this paper is to estimate the thermal efficiency, capital investment and total cost of indirect coal liquefaction processes. Over the past five years many study designs have been performed on the production of methanol and other indirect liquids from coal. Some of these are original designs, while others are secondary studies, taking one or more original designs and adjusting economic parameters, etc. Figure 1 shows the available indirect coal liquefaction studies and their chronology and interrelationship. Since the secondary studies only modified the economic basis of the original studies referenced and not the basic design and each secondary study used a different basis, preventing intercomparison, this study will restrict itself to the original studies and attempt to place them all on one single comparable basis. The following is a list of these original studies:

"Screening Evaluations: Synthetic Liquid Fuels Manufacture," Ralph M. Parsons Company for EPRI, August, 1977, EPRI AF-523.[1] (This report estimates the cost of methanol from four different gasification technologies, Foster-Wheeler, BGC-Lurgi, Koppers-Totzek, and Texaco, with Chem Systems methanol synthesis. The study also looks at the Fischer-Tropsch process following BGC-Lurgi gasification.)

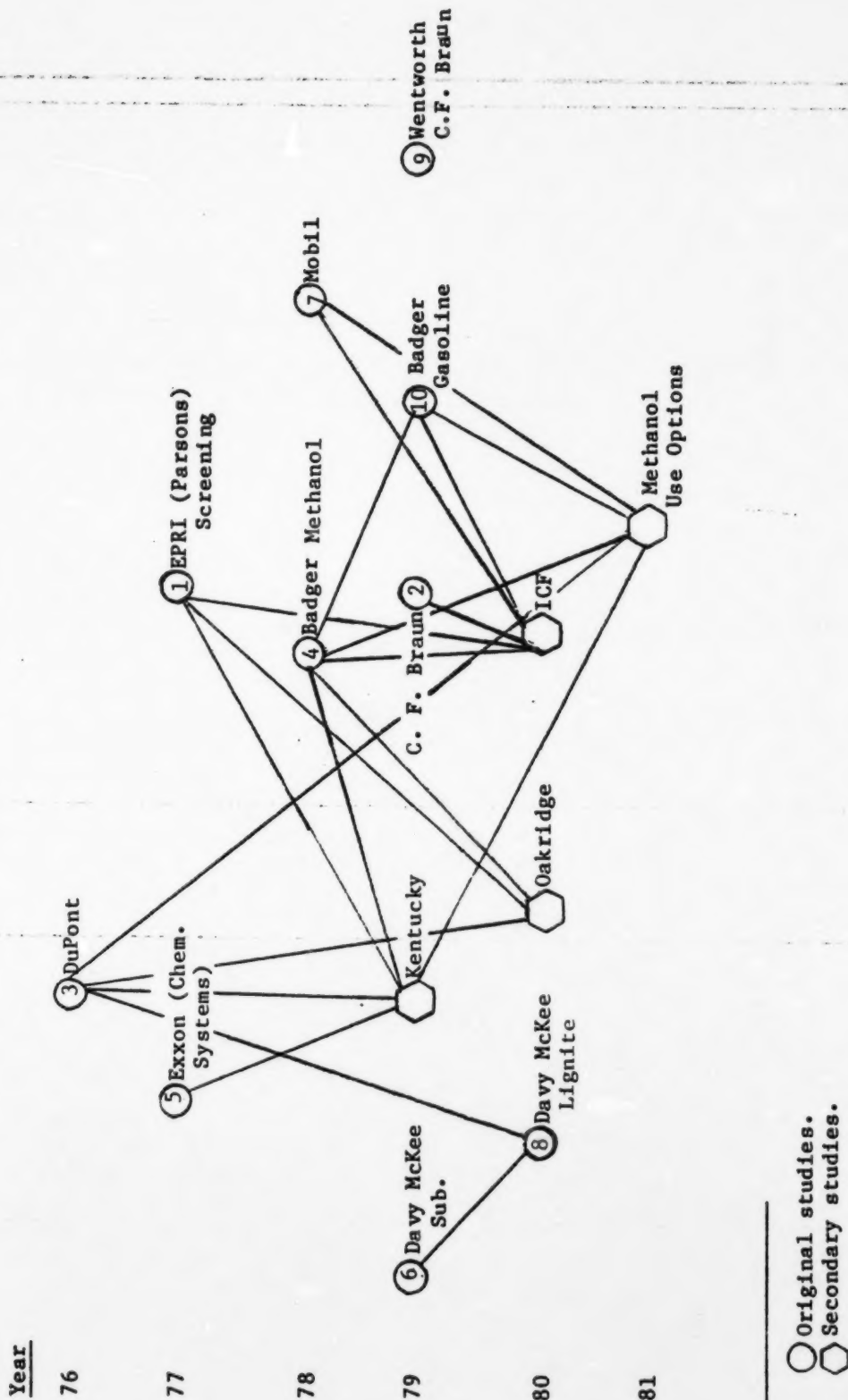
"Coal to Methanol Via New Processes Under Development: An Engineering and Economic Evaluation," C.F. Braun and Company for EPRI, October, 1979, EPRI AF-1227.[2] (This report covers two coal to methanol processes: Illinois No. 6 coal to methanol via Texaco gasification and Chem Systems methanol synthesis, and Wyodak coal to distillate fuel and vacuum residual oil via a non-catalytic hydroliquefaction process in which the residual oil is processed into methanol by the same process as the coal.)

"Economic Feasibility Study, Fuel Grade Methanol From Coal For Office of Commercialization of the Energy Research and Development Administration," McGeorge, Arthur, DuPont Company, for ERDA, 1976, TID-27606.[3] (Eastern coal to methanol via Texaco gasification with ICI synthesis.)

"Conceptual Design of a Coal-To-Methanol Commercial Plant" (Vols. I-IV), Badger Plants, Inc., for DOE, February, 1978, FE-2416-24.[4] (Eastern coal-to-methanol via Lurgi "slag-bath" gasification and Lurgi low pressure methanol synthesis technology.)

"Production Economics for Hydrogen, Ammonia, and Methanol During the 1980-2000 Period," Cornell, H.G., Heinzelmann, F.J., and Nicholson, E.W.S., Exxon Research and Engineering Co., April, 1977.[5] (Eastern coal to methanol via Koppers-Totzek and Shell-Koppers gasification with ICI synthesis.)

Figure 1
Methanol Report "Tree"



"Methanol From Coal, An Adaptation from the Past," E.E. Bailey, (Davy McKee), presented at The Sixth Annual International Conference; Coal Gasification, Liquefaction and Conversion to Electricity, University of Pittsburgh, 1979.[6] (Subbituminous coal to methanol via Winkler gasification and ICI synthesis.)

"Research Guidance Studies to Assess Gasoline From Coal By Methanol-To-Gasoline and Sasol-Type Fischer-Tropsch Technologies - Final Report," Schreiner, Max, Mobile R&D Company, for DOE, August, 1978, FE-2447-13.[7] (Comparison of eastern coal to methanol and SNG, and gasoline and SNG by Lurgi gasification/Lurgi synthesis/Mobil MTG with gasoline from Lurgi gasification and Fischer-Tropsch synthesis.)

"Lignite-to-Methanol: An Engineering Evaluation of Winkler Gasification and ICI Methanol Synthesis Route," DM International, Inc. for EPRI, October 1980, EPRI AP-1592, Project 832-3.[8] (Lignite to methanol via modified Winkler and ICI synthesis.)

"Production of Methanol from Lignite," Wentworth Bros., Inc., and C.F. Braun and Co., for EPRI, September 1979, EPRI AF-1161, TPS-77-729.[9] (Lignite to methanol via Texaco gasification and WBI synthesis.)

"Conceptual Design of a Coal-to-Methanol-to-Gasoline Commercial Plant," for DOE, March 1979, FE-2416-43.[10] (Adds Mobil process to methanol design of study no. 4 above.)

Methanol: To estimate the cost of producing methanol, all of the design studies were: 1) normalized to a production yield of 50,000 fuel oil equivalent barrels per calendar day (FOEB/CD) and a common financial basis and 2) inflated to \$1981, as discussed in a previous report.[10a] Of the thirteen designs contained in the above ten studies, nine used bituminous coal, two used subbituminous coal and two used lignite as a feedstock. The studies included eight different coal gasification technologies (Foster-Wheeler (1), BGC/Lurgi (1), Koppers-Totzek (2), Texaco (4), Lurgi (1), "Slag-Bath" (1), modified Winkler (2) and Koppers-Shell(1)) and four different types of methanol synthesis processes (Lurgi (2), ICI (5), Chem Systems (5), and Wentworth Bros. (1)). The Winkler, Lurgi and Koppers-Totzek gasifiers have been proven on a commercial scale and the Texaco process is very close to commercialization. Of the synthesis units, ICI and Lurgi are used extensively today. Wentworth Bros. claim that their process is commercial and Chem Systems is a new process which is still being tested.[21] Lurgi and ICI have been competing for the last ten years and both have highly developed processes, good efficiencies and, according to Parsons,[1] room for further improvement is small. In addition, Parsons states that the Chem Systems process only shows a slightly higher thermal efficiency and lower capital cost than the ICI system. Since the costs of the proven ICI and Lurgi synthesis processes are indistinguishable

and it appears that the cost for the Chem Systems process is only slightly lower, it has been decided to place most of the emphasis here on the effect of the various gasification technologies costs which appear to be more significant.

The original ranges of product costs and capital costs reported by the thirteen studies are very large due at least in part to the large range in plant size (\$3.74-12.55 per mBtu for product cost and \$0.401-\$5.05 billion for capital, \$1981, for plants ranging from 2,000-58,000 ton per day of methanol). With this type of data it is very difficult to estimate the actual cost of methanol, let alone compare it with any other coal technologies. After normalizing the costs for the thirteen studies the ranges of costs are much smaller.

For bituminous coals the product cost ranged from \$4.65-9.05 per mBtu for the low capital charge rate (CCR) and \$8.14-12.54 per mBtu for the high CCR. The gasifiers used in these studies are Foster-Wheeler, BGC-Lurgi, Koppers-Totzek, Lurgi Slag Bath, and Texaco(2).

Of these gasifiers the Koppers-Totzek is proven, and the remainder represent advanced technology. The cost of methanol from these gasifiers are presented in Table 1. When using the Koppers-Totzek gasifier the cost ranges from \$7.23-12.42/mBtu depending on the capital charge rate; for the Texaco gasifier the cost ranges from \$5.90-6.48 and \$9.44-10.41/mBtu; for the other advanced technology the cost ranges from \$5.30-6.08 to \$8.74-9.78/mBtu.

Both the Texaco and Koppers-Totzek gasifiers are entrained bed units which seems to emphasize the statement that entrained bed gasifiers are the only commercially-available (or near-commercial) reactors today which can economically gasify eastern bituminous coals (eight of the nine cases investigated used entrained bed gasifiers).

The range of instantaneous plant investment for the nine cases was \$1.93-\$2.92 billion (50,000 FOEB/CD plant). As shown in Table 1, the instantaneous plant investment for the methanol plant using bituminous coal ranged from \$1.99 to \$2.21 billion when the Texaco gasifier was used, \$2.92 when the Koppers-Totzek gasifier was used, and ranged from \$1.93-2.22 billion when the other advanced technology gasifiers were used.

The range of product and capital costs for methanol from subbituminous coals and lignite are smaller than that of bituminous. Of the two studies using subbituminous coals, one uses proven gasification and synthesis technology, Lurgi/Lurgi [7], while the other uses a gasification technology which the manufacturer claims is "here now," and a proven synthesis process, modified Winkler/ICI.[6] The average product cost range is fairly

Table 1

Product and Capital Costs of Selected
Coal Liquefaction Processes(1981 Dollars)

<u>Indirect Liquefaction</u>	<u>Product Mix</u>	<u>Product Cost (\$/mBtu)</u>		<u>Capital Cost** (Billions of Dollars)</u>
		<u>11.5% CCR</u>	<u>30% CCR</u>	
<u>Process</u>				
Texaco (Bituminous)[2,3]	100% MeOH*	5.90-6.48	9.44-10.41	1.99-2.21
Koppers (Bitum.)[1]	100% MeOH*	7.23	12.42	2.92
Advanced Technology (Bituminous)[1,4]	100% MeOH*	5.30-6.08	8.74-9.78	1.93-2.22
Lurgi (Subbit.)[7]	47.9% MeOH*	7.04	12.48	2.59
	49.7% SNG	5.63	9.98	
	2.4% Gasoline	7.04	12.48	
Modified Winkler (Lignite)[8]	100% MeOH*	5.70	9.56	2.17
Texaco (Lignite)	100% MeOH*	6.92	12.24	3.00
Lurgi-Mobil MTG (Subbit.)[7]	41.2% Reg. Gasoline	8.01	14.35	2.95
	53.3% SNG	6.41	11.48	
	5.5% LPG	6.25	11.20	
Mobil MTG Incremental Cost	85-90% Reg. Gasoline	1.45	2.87	0.68
	10-15% LPG			
Fischer Tropsch[7]	1.8% LPG	6.56	11.36	3.00
	64.5% SNG	6.82	11.80	
	2.6% Alcohols	8.52	14.75	
	25.3% Gasoline	8.52	14.75	
	4.6% Diesel Fuel	7.67	13.28	
	1.3% Heavy Fuel Oil	6.56	11.36	

* MeOH = 95-98% methanol, 1-3% water, and the remainder higher alcohols.

** Instantaneous capital costs.

small, \$6.16-\$6.34 per mBtu for the low CCR and \$10.26-\$11.24 per mBtu for the high CCR. The instantaneous plant investment range is \$2.10-\$2.59 billion. Although the costs seem to compare favorably, only the Lurgi/Mobil prices are shown in Table 1. This is because the modified Winkler/ICI plant size had to be scaled up significantly where as the Lurgi/Mobil plant size was much closer to the selected 50,000 FOEB/CD and was therefore considered more accurate.

Four lignite cases were studied. However, two of these cases used Texaco gasifiers with coal slurry concentrations which are still in a developmental stage. The other cases involved the Texaco gasifier (at an appropriate coal slurry concentration) and the Winkler gasifier. At this slurry concentration the Texaco gasifier appeared to have a large economic disadvantage relative to the Winkler gasifier, so the Winkler was chosen as the best design. The resulting product costs for the low and high CCRs are \$5.70 and \$9.56/mBtu, respectively. The instantaneous investment plant cost is \$2.17 billion. These costs are shown in Table 1.

In summary, the prices which have been chosen for this study represent two commercially proven gasification technologies: Koppers-Totzek and Lurgi, a modified Winkler, for which its manufacturer will back financially, and the near commercial Texaco gasifier. For bituminous coals, the Koppers-Totzek prices are higher than Texaco's because the former operates at atmospheric pressure.

MTG: To evaluate the cost of producing gasoline from coal utilizing Mobil's methanol-to-gasoline (MTG) process, two different studies by Mobil and Badger were analyzed in the same manner as the methanol studies.[7,10] Gasoline costs from these two studies varied widely. Therefore, it was preassumed that incremental product and capital costs for Mobil's MTG gasoline relative to methanol could be determined from both studies and be more accurate since methanol costs (capital and product) were available for the same technology by the same designers.[7,4] When the cost of gasoline was compared to that of methanol, the incremental cost of gasoline for both studies was very close. Since the MTG process is a patent of Mobil's, it is believed that their study is more reliable; therefore their costs were used in preference to Badger's, which were slightly higher.

The Mobil study analyzed a few different cases with respect to the Mobil MTG process. The most economical was the case which produced gasoline and SNG as the major products. For this case, the average product cost ranged from \$7.06-12.65 depending on the CCR. The total instantaneous plant investment was \$2.95 billion. These costs are shown in Table 1. By comparing this case with Mobil's other case (methanol from Lurgi gasification of subbituminous coal) an incremental cost of gasoline relative to methanol was determined. Based on a 50,000 FOEB/CD MTG unit, the incre-

mental cost of gasoline over methanol was found to range from \$1.45-2.87 per mBtu depending on the CCR. The incremental instantaneous investment was found to be approximately \$680 million for a plant producing all methanol and then gasoline. These costs are also shown in Table 1.

Fischer-Tropsch: There were two studies which investigated Fischer-Tropsch synthesis technology, Parsons and Mobil.[1,7] Since the Mobil study was based on a more thorough design than the Parsons study, its costs were used in preference. The instantaneous plant investment cost for the Mobil case was \$3.00 billion. Its average product cost ranged from \$7.60-13.38 per mBtu depending on the CCR. The costs of the products from this case are presented in Table 1. The Mobil study was also used to determine the average product cost difference between Fischer-Tropsch synthesis and methanol synthesis plants. The instantaneous plant investment difference is \$355 million and the operating cost difference is \$67 million with the Fischer-Tropsch case costing more. The figures translate into an average product cost difference of \$1.00/mBtu.

II. History of Methanol Production

The primary source of all methanol prior to the 1920's was the destructive distillation of wood. In this pyrolysis process air was excluded while the wood was heated to a temperature of 160-400 degrees Centigrade. As the components of the wood heated they volatilized and thermally decomposed. The products were separated into gases and a condensed liquid called pyroligneous acid. Upon further distillation this liquid could be separated into acetic acid, acetone and rather impure methanol. Since the yield was three to six gallons per ton of wood, the product was very expensive.[12]

During the pre-World War I period, the development of a synthetic methanol process began in Germany and France. Between 1910 and 1916 there were several patents issued in Europe describing the chemical reaction of carbon monoxide (CO) and hydrogen (H₂) to form alcohols, ketones, aldehydes, etc. The reaction was carried out at temperatures of 300 to 400 degrees Centigrade and at pressures at or above 1500 psi. Catalysts containing chromium, zinc, manganese and cobalt, or their oxides were used to help the conversion of the carbon monoxide and hydrogen to methanol.[12]

In 1923, BASF in Germany became the first company to produce commercial-scale synthetic methanol. The U.S. started importing synthetic methanol produced from coal or coke in 1924. Soon Commercial Solvents Corporation and DuPont became interested and by 1928 each had a commercial plant producing methanol in the U.S.[12]

When coal and coke derived synthetic methanol hit the U.S. market there was an enormous difference between the price of natural and synthetic methanol. Natural methanol cost 68 cents per gallon while synthetic methanol could be made for 36 cents per gallon. The price competition was so great that the natural wood distillers united and managed to persuade the tariff commission to increase the import tariff to 18 cents per gallon. They also were able to get legislation passed which mandated the use of natural methanol to denature ethanol, thus securing a third of the total methanol market.[12]

The wood distillers managed to keep the price of natural methanol competitive for a number of years through consolidation and larger more efficient plants. However with the large discoveries of petroleum and natural gas and the mass production of high-purity methanol, the synthetic manufacturers were soon able to lower the price of methanol beyond reach, leaving natural methanol producers to their captive denaturant market.[23]

The first plants were built in conjunction with other plants to make use of carbon dioxide or hydrogen by-products. However as the demand for methanol grew, plants were built specifically for methanol production. The first feedstock to be gasified to carbon monoxide and hydrogen was coal. Later the feedstock was shifted to oil and then natural gas as large discoveries of these sources were made and their prices dropped. Natural gas was an ideal feedstock because it contained very little, if any, sulfur and its price was very low. Thus by the 1960's, synthetic methanol in the U.S. was almost entirely produced from natural gas utilizing the high-pressure methanol synthesis process.[23]

By 1967 the combination of a common feedstock, comparative technology and a competitive market had stabilized the price of methanol at 27 cents per gallon. However, in 1967, Imperial Chemical Industries (ICI) introduced a newly developed low-pressure synthesis process based on a copper-zinc-chromium catalyst in place of the zinc-chromium catalysts previously used. Since these copper catalysts were more reactive than the others, lower operating pressures and temperatures could be used. In fact, by the latter part of 1971, the selling price of methanol had dropped to 11 cents per gallon.[23]

In the 70's the increasing cost of production, the demand for low-sulfur natural gas and the OPEC oil embargo of 1973 brought into focus the energy crisis and the finite supply of fossil fuels. The tripling of oil prices and doubling of the cost of natural gas caused the price of methanol to triple between 1973 and 1975 (14 cents/gal to 42 cent/gal). Since 1975 the price of methanol has continued to increase with that of natural gas. The current price for methanol is between 72 and 80 cents per gallon.[23]

Between the time it was first introduced into the U.S. and today, methanol has exhibited a dramatic growth. For the first 45 years there was 13.7 percent average annual growth rate.[12] In the 1930's plant sizes ranged from 20 to 40 tons of methanol per day. By the early 50's the size has risen to 150-200 tons/day. In the 70's the capacity has gone from 1,500 to 2,000 tons/day to single trains of 5,000 tons/day.[23]

Methanol production in the United States is now near 4 million tons per year or about 70,000 barrels per day (BPD). Virtually all of this is produced from natural gas.[12] The natural gas (essentially pure methane) is reformed with steam to produce a synthesis gas consisting mainly of carbon monoxide and hydrogen. After purification, the synthesis gas is compressed and combined in a catalytic converter to produce methanol. The reaction is highly exothermic while the conversion per pass is relatively small (2-10 percent). Large volumes of unconverted gas are recycled through the converter in order to achieve high overall conversion and to assist in removing the exothermic heat of reaction. Overall CO conversions of 96 to 99 percent can be obtained.[13]

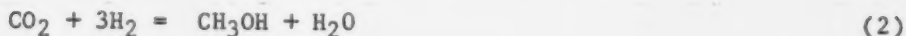
III. The Methanol Production Process

The basis of all processes for manufacturing synthetic methanol is the catalytic reaction of carbon monoxide and carbon dioxide with hydrogen to produce methanol.[12] These reactions are shown below.

Carbon Monoxide + Hydrogen = Methanol



Carbon Dioxide + Hydrogen = Methanol + Water



The source of carbon monoxide or carbon dioxide is usually derived from the partial combustion of a hydrocarbon fuel such as coal, coke, natural gas, naptha, or a heavy petroleum fraction.

The primary source of hydrogen is water and the hydrogen contained in the feedstock, which is the case of coal is very low (3-6 percent). The reactions shown in Equations (1) and (2) are carried out at pressures between 750 and 4500 psi at a temperature of 250 to 350 degrees Centigrade in the presence of a metal or metal oxide catalyst. The metals used depend upon the process, and are usually proprietary. Catalysts may contain zinc, chromium, or copper-based compounds or oxides.

A description of a typical coal to methanol process follows.[11]

Coal Receiving and Preparation: Vibrating feeders transport the coal to the sizing equipment, ring mill crushers and rod mills where the coal is sized for the specific gasifier in which it will be processed.

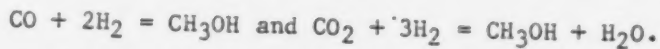
Coal Gasification: The coal is heated to very high temperatures and partially-oxidized to carbon monoxide and hydrogen in the presence of oxygen (or air) and steam. The majority of the sulfur is converted to hydrogen sulfide with some production of carbonyl sulfide. The nitrogen in the coal is converted to free nitrogen combined with some traces of ammonia and hydrogen cyanide. The ash is removed from the bottom in a dry or molten slag depending on the temperature and gasification technique used.

Gas Cooling: The hot raw gas is cooled and scrubbed with recycle gas liquor or sour water from the shift converter. Then the gas is cooled further in a heat exchanger where steam is produced by the waste heat.

Gas Shift: Here the ratio of hydrogen to carbon monoxide is increased by adding steam and pushing the following water-gas shift reaction to the right: $\text{CO} + \text{H}_2\text{O} = \text{CO}_2 + \text{H}_2$.

Acid Gas Removal: In this process the sulfur is removed from the synthesis gas to prevent poisoning of the methanol synthesis catalyst. In the Selexol process hydrogen sulfide is removed first, and then carbon dioxide and carbonyl sulfide are removed. In the following Rectisol process, naphtha, HCN and water are removed by washing the gas with a small quantity of methanol.

Methanol Synthesis: In this stage clean shifted gas is catalytically converted into crude methanol by the following two reactions:



Auxiliary Facilities: The functional relationships of the auxiliary facilities to the major process areas are as follows:

Water Supply - provides for treatment, storage and distribution of process water requirements, including makeup cooling water.

Water Cooling - provides for treatment, storage and distribution of process cooling water.

Oxygen Production - cryogenically separates air into oxygen and nitrogen. Oxygen is used in coal gasification. Some of the nitrogen is used in carbon dioxide removal, the remainder being vented to the atmosphere.

Slag Removal/Char Recovery - separates dissolved gases from the raw gasifier slag. Also separates char from the slurry produced in gas cooling. Overhead gases and char are used to generate steam. Wet slag is sent to slag/ash disposal.

Slag/Ash Disposal - combines wet slag, bottom ash from steam generation and dusty liquor from flue gas cleanup. Dewatered slag/ash mixture is suitable for landfill disposal. Wastewater is used for preparation of coal slurry for coal gasification.

Sulfur Recovery - processes the acid gas stream produced during hydrogen sulfide removal, converting H_2S to elemental sulfur. Process technology employed is usually Claus (bypass type configuration).

Steam Production - uses char, purge and overhead gases, along with supplemental coal to provide plant steam/power requirements.

Flue Gas Clean-up - renders steam generation product gases environmentally acceptable for stack discharge to atmosphere.

IV. Gasification Technology

The gasification of coal began in the early 1850's when it was discovered that the gas could be burned more efficiently than solid coal and it was cleaner and easier to use. The technology developed fast and by the 1850's gas light for streets in London was commonplace. Between 1935 and 1960 there were close to 1,200 municipal "gasworks" serving larger towns and cities in the U.S. However the introduction of natural gas pipelines in the 1930's initiated the decline and almost disappearance of coal gasification in the U.S. With the increased cost of natural gas, interest in coal gasification has been renewed.

Numerous processes are now being developed to gasify coal, the most abundant hydrocarbon resource in the U. S., to low-, medium- and high-BTU gas. In the gasification process coal is reacted with a mixture of steam and air or steam and oxygen. With the former, a low-BTU gas is produced with a heating value between 100 to 200 BTU/scf.[15] This low BTU gas is made up of nitrogen, carbon monoxide, hydrogen, carbon dioxide and water.[22] This gas has a low BTU content because it contains a large portion of nitrogen (since air contains 80 percent nitrogen) which dilutes the energy content of the carbon monoxide and hydrogen produced. If the coal is mixed with steam and oxygen a medium BTU gas is produced consisting of carbon monoxide, hydrogen, carbon dioxide and methane, which has a heating value between 250 and 400 BTU/scf.[15] High BTU gas (or synthetic natural gas (SNG)) with a heating value of 970 BTU/scf can be produced from medium BTU gas by methanation or hydrogen removal.[29,30]

There are several factors such as thermal efficiency, reliability, capital investment, coal flexibility and product spectrum

which are important and should be considered when comparing gasifiers. Table 2 shows some of these comparisons for different gasification systems. For instance thermal efficiency is important from a processing view. To achieve maximum efficiency a gasifier should have low oxygen and steam demands, low unburned carbon and heat losses, and should operate at high temperatures.[14] However, since some of these factors are not compatible with others, it is almost impossible to obtain a gasifier which optimizes each factor. For example high oxygen requirements (lower efficiency) are necessary to obtain high carbon conversions (higher efficiency) and to avoid large by-product formations (lower capital investment).[14] In addition, elevated pressures (high efficiency) produce more by-products[15] and interfere with reliability[14] but reduce raw gas compression requirements. Other desirable factors are high temperatures which reduce by-products and increase coal flexibility and capacity.[14] In general then, it is apparent that there is a trade off between efficiency and some of these other factors. The best gasifier will be that process unit which optimizes the majority of these factors while achieving the highest efficiency.

Before discussing the individual gasifier types it is helpful to examine the properties of coals used in the U.S. There are four properties of coal which are important in the process selection of gasifiers: 1) ash fusion temperature, 2) free swelling index (FSI), 3) moisture, and 4) sulfur content. The ash fusion temperature is that temperature at which the ash becomes fluid. FSI is a measure of a coal's tendency to agglomerate or cake when heated; the higher the FSI, the greater the agglomeration.

Eastern coals (predominantly bituminous) typically have low fusion temperatures (1990-2200°F), moderate to high FSI, low moisture (4-10 percent by weight as received) and high sulfur (3-5 weight percent). Western coals (mainly subbituminous) exhibit high fusion temperatures (2300-2400°F), low FSI, high moisture (28 weight percent) and low sulfur (0.3-0.5 weight percent). Lignite, which is found predominantly in North Dakota, has an even higher moisture content (35 weight percent) and a lower percentage of sulfur (0.2 weight percent).

Coal gasifiers are classified according to the way coal is fed to them. The three main gasifier categories are the fixed or slow moving bed, the fluidized bed and the entrained bed. Tables 3 and 4 show a list of coal gasifiers by type and Table 5 summarizes the advantages and disadvantages of the three different types of gasifiers.

A. Fixed or Slow Moving Bed

Fixed or slow moving bed gasifiers consist of beds that carry or move the coal vertically downward through the zone where it is heated and decomposed. These gasifiers can be further divided

Table 2

Comparison of Gasification Systems

<u>Name</u>	<u>Lurgi</u>	<u>BGC-Lurgi</u>	<u>Winkler</u>	<u>Koppers- Totzek</u>	<u>Texaco</u>	<u>Shell- Koppers</u>
Bed Type	Fixed	Fixed	Fluid	Entrained	Entrained	Entrained
Commercial	Yes	Near	Yes	Yes	Near	Near
Coal Flexibility	Western	Western	Western	All	All	All
By-Product	Yes	Yes	No	No	No	No
Efficiency(%)	64	72	57	58	68-72	75
Capacity (STPD Coal)	500	1,250	1,000	850	2000	1,000

Table 3

Coal Gasifiers

	<u>Pressure</u>	<u>Oxygen or Air</u>	<u>Agglomeration Prevention</u>
<u>FIXED OR SLOW MOVING BED</u>			
	<u>Dry Ash, Single Stage</u>		
Gegas	To 500 psig	Air	Stirrer paddles
Lurgi	To 450 psig	Oxygen or air	Rotating blades
Merc	To 105 psig	Air	Spiraling stirrer
Riley-Morgan	40 in water	Air	Agitator in rotating bed
Wellman-Galusha	10 in. water	Air	Spiraling arms
Wilputte	atm	Air	Rotating arm
	<u>Dry Ash, 2-Stage</u>		
ATC/Wellman	atm	Air	None
FW/Stoic	atm		None
Ruhr-100	1,500 psig	Oxygen	Stirrer blades
Woodall-Duckham	40 in. water	Air or oxygen	None
	<u>Slagging</u>		
BGC/Lurgi	To 400 psig	Oxygen	Stirrer
Gferc	To 400 psig	Oxygen	Stirrer
<u>FLUIDIZED BED</u>			
Winkler	atm	Oxygen or air	
Rheinbraun	150 psig	Oxygen	
CO ₂ acceptor	150 psig	Air	
Hygas	1,200 psig	Oxygen or air	
Synthane	1,000 psig	Oxygen	
Westinghouse	225 psig	Air	
U-gas	350 psig	Oxygen or air	
Cogas	10 psig	Air	
EDS (Exxon)	500 psig	None	
<u>ENTRAINED FLOW</u>			
	<u>Single-Stage</u>		
Bell HMF	to 225 psig	Air	
Koppers-Totzek	10-12 psig	Oxygen and steam	
Mountain fuel	to 150 psi	Oxygen and steam	
Shell-Koppers	to 450 psi	Oxygen or air	
Texaco	to 1,200 psig	Oxygen or air	
Bi-gas	To 1,500 psig	Syngas and steam	Oxygen and steam air
C-E	atm	Syn gas and air	Air and steam
Foster Wheeler	atm	Syn gas and steam	Air and steam
Peatgas	To 500 psig	Syngas	Oxygen and steam
Rockwell Int'l	To 1,500 psig	Hydrogen	Not known

Source: Institute of Gas Technology[29], Oil and Gas Journal, 7/16/81, pg. 57

Table 4

Coal Gasification Process Technology Status

<u>Gasifier</u>	<u>Technology</u>	<u>Process Status</u>	<u>Location</u>
<u>Fixed Bed</u>			
Lurgi, Dry Ash	1st Generation	Commercial	Worldwide
British Gas/Lurgi Slagging	2nd Generation	Semicommercial	Westfield, Scotland
Wellman Galusha	1st Generation	Commercial	14 operating in U.S. others outside U.S.
KilnGas	2nd Generation	Pilot (1971-)	Oak Creek, Wis.
<u>Fluidized Bed</u>			
Winkler	1st Generation	Commercial	Worldwide
U-GAS	2nd Generation	Pilot (1974-)	Chicago, IL
Westinghouse	2nd Generation	PDU (1975-)	Waltz Mill, PA
<u>Entrained Flow</u>			
Koppers-Totzek	1st Generation	Commercial	Worldwide
Texaco	2nd Generation	Pilot	Montebello, CA Mussel Shoals, Ala. & outside the U.S.
Combustion Engineering	2nd Generation	PDU (1974-)	Windsor, Conn.
Shell-Koppers	2nd Generation	Pilot	W. Germany

[15] Source: Oil & Gas Journal; June 29, 1981, pg. 106.

Table 5

Gasifier Characteristics

<u>Fixed Bed Gasifier</u>	<u>Fluidized Bed Gasifier</u>	<u>Entrained Bed Gasifier</u>
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Advantages

Extensive practical experience	Uniform temperature and compositions throughout fluidized zone	Highest capacity per unit volume
High carbon conversion efficiency	Excellent solid-gas contact	Produces inert slagged ash with low carbon content
Low temperature operation	No internal moving parts	Product gas free of tars and phenols
Large fuel inventory provides safety, reliability and stability	Can handle wide variety of coals Large fuel inventory provides safety, reliability, and stability	Handles all types of coal No moving parts and has simple geometry

Limitations

Sized Coal required	Distributor plate design is critical	Less developed than fixed bed
Coal fines must be disposed of or handled separately	Requires dry coal for feeding	Critical design areas include combustor nozzles and heat recovery in presence of molten slag
Product gas contains tars and heavier hydrocarbons		Smallest fuel inventory; requires advanced control techniques to ensure safe reliable operation
Lowest capacity due to limited gas-flow rates	Removal of fines required to prevent elutriation or flow instability	
Internal moving parts with higher degree of mechanical complexity	Fluidization requirements sensitive to coal characteristics	
Caking coal technology not commercially proven		

into two types which describe the flow of air in them: updraft and downdraft.

The simplest air gasifier is the updraft or countercurrent gasifier which introduces air at the bottom of the furnace where it first comes into contact with the hottest temperatures of the reactor. Since the combustion gases immediately enter a zone of excess char, any carbon dioxide or water present is reduced to carbon monoxide and hydrogen by the excess carbon. In addition to producing the desired products, carbon monoxide and hydrogen, these hot gases contain large amounts of tars, phenols, cresols and other oxygen containing organic compounds. As the gas rises its temperature decreases as heat is transferred from the hot gas to the cooler incoming coal. This low temperature hinders the oxidation of the coal and is the major cause of the by-products produced.[22]

One of the problems caused by these by-products (chemicals, oils and tars), is that they condense in the cooler regions, causing maintenance problems. In addition, these components contribute to the majority of environmental problems associated with fixed bed gasification systems.[18]

The downdraft gasifier is specifically designed to eliminate the tars and oils associated with the updraft gasifier. Tars and oils are formed near the middle of the bed (where air is injected) and carried by the airflow through a relatively large hot zone in which they have time to further decompose or be cracked to simpler gases or char. One of the important results of this cracking is that an effect called "flame stabilization" occurs which maintains the temperature range between 800°C to 1000°C. When the temperature rises, endothermic reactions predominate, causing the gas to cool; when the temperature drops, the exothermic reactions predominate, thus heating the gas.[22]

The tars and oils are reduced to less than 10 percent of the amount produced in updraft gasifiers thereby making gas clean-up easier and less expensive. Since the gas velocities are low in both updraft and downdraft gasifiers, the ash settles through the grate so that very little is carried with the gas.[22]

One example of a moving bed gasifier is the Lurgi gasifier which is commercially available through Lurgi Kohle and Mineraloeltechnik. In the Lurgi process, coal is fed into the gasifier via automatically operated coal locks. As the bed of coal moves from the top to the bottom of the gasifier it comes in contact with a counter-current hot gaseous mixture of oxygen and steam introduced at the bottom which successively dries, devolatilizes and gasifies the coal. The partial oxidation of the coal with oxygen supplies the necessary heat for the coal gasification while the addition of steam prevents the temperature from rising above the ash fusion (or melting) point. The ash left after gasification is removed by a rotating grate at the bottom of the gasifier.

As shown in Table 3, the Lurgi "dry-ash" fixed bed gasifier is a first generation unit which has been commercially proven and is used worldwide.[14,15] The Sasol I plant in South Africa which has been operating for over 25 years utilizes the Lurgi gasifier (and also the Fischer-Tropsch synthesis unit) to produce 10,000 barrels per day of fuel. (It and Sasol II are the only existing commercial-size coal-liquefaction plants in the world.) The Dunn Nakota project, which is hoping to produce 85,000 barrels per day of methanol by 1987 via Lurgi gasification, could be the largest commercial-scale coal gasification process built in the U.S.[31] The main disadvantages of the Lurgi gasifier are that it 1) has problems with the caking eastern coals, 2) produces byproducts, 3) has high steam requirements and 4) has a low capacity per volume of gasifier (i.e., high capital cost).

The BGC/Lurgi slagging gasifier is a second generation reactor which is now being tested in Scotland by Lurgi and British Gas Corp with support from 13 U.S. companies and DOE.[30] In this gasifier, coal is fed into the top of the unit by a distribution system. As the coal descends in a moving bed, it is successively dried, devolatilized and gasified. At the bottom of the gasifier oxygen and steam are fed and slag is withdrawn. The operating pressure is 300-350 psig with a gas temperature of 800-1100°F and an ash temperature over 2000°F so the slag can be removed in a molten form.[1] Because it does operate in the slagging mode it can tolerate a higher throughput of coal and oxygen without entraining coal dust in the product gas.[16]

The latest papers describe this technology as near commercial.[14,15] Its improvements over the older Lurgi dry-ash gasifier are a higher efficiency and a reduction in steam use. However, it still has problems with caking eastern coals and still produces by-products.

B. Fluidized Beds

Over the last 60 years fluidized beds have been developed to provide uniform temperatures and efficient contact between gases and solids. This is accomplished by blowing gas upward through a bed of solid coal so rapidly that the bed becomes suspended and churns as if it were a fluid. Fluidized reactors are more compact because they have a higher throughput (due to higher reaction rates), but the higher velocity of the gas carries out ash and char with it that must be removed by cleaning the product gas. The fluid bed often contains limestone to react with and remove the sulfur from the coal. Fluidized bed reactors have a considerably faster heating rate than moving bed gasifiers and, therefore, the reactor temperature must be held below the softening or initial deformation temperature of the coal ash which is typically well below 1040°C. However, at this temperature many undesirable by-products are stable and the churning of the bed enables materials at all stages of decomposition to be found

throughout the bed. Because of this contact, tars and oils have a tendency to escape from the heating zone before they can be fully decomposed. This removal and disposal of these by-products can pose a number of environmental problems for the fluidized reactor. Claims that they can produce very low tars and char with recirculation still remain to be proven.[22] Since typical operating temperatures are low with respect to the ash melting temperature of coals, the fluid-bed gasifier also has problems with eastern coals.

The Winkler fluid-bed gasifier is a first generation unit which is commercially proven and used around the world.[15] According to DM International over 70 Winkler gasifier have been built.[8] The two main disadvantages with the Winkler are that it operates at atmospheric pressure (large volume per throughput) and that it has a tendency to clog when using eastern coals. A pressurized modification of the Winkler is now under development which should improve its efficiency.[8,14]

In the two designs of lignite gasification that will be reviewed in this study, modified Winkler gasifiers have been used. In both cases the modified Winkler operates at a higher pressure (65 psig) than the established Winkler which operates at atmospheric pressure (14.7 psig). The lignite is dried from 35 percent moisture to 8 percent and is then continuously fed by a pressure lock and screw conveyor system into the Winkler gasifier where it is maintained as a fluid bed at 65 psig. Steam is injected near the bottom of the reactor to fluidize the coal and to cool the larger ash particles discharging from the gasifier bottom while steam and oxygen are injected at several points within the bed to gasify the coal. Since the gasifier operates at high temperatures (1800-1900°F), tars, oils, gaseous hydrocarbons and carbon present are converted to carbon oxides and hydrogen. Only a small percentage of methane is left in the raw gas product. In the fluidized bed, heavier particles such as ash fall down through the bed into the char discharge, while lighter particles are carried out of the bed by the product gas. In the Winkler gasifier approximately 70 percent of the total char is entrained in the hot product gases leaving the top of the reactor.

This modified Winkler is still being tested and therefore it cannot be considered to be commercially proven. However, since Davy McKee believes that this design contains equipment similar to other high pressure units, they feel that the gasifier is feasible and are therefore prepared to offer commercial guarantees.

C. Entrained Bed

The entrained bed gasifier, which dates back to the 1950's, is the most recently developed gasifier. In the entrained bed gasifier fine particles of coal are suspended in a stream of oxygen which moves rapidly into and through the decomposition

zone. The entrained bed gasifier is typically operated at a temperature above the melting point of the coal ash. At this temperature, which is typically 1260-1316°C, the gasification reaction rates are much faster and many of the undesirable by-products associated with the fixed bed and the fluid bed systems are unstable and are destroyed. When the entrained bed gasifier is operated at pressures substantially above atmospheric, high throughput and high single pass conversion can be obtained. One drawback is that the feedstock must be reduced to a relatively small size which would add to the total preparation cost. However, there is a tradeoff since the smaller particles are more efficiently gasified.

These gasifiers are also called "slagging" because they remove the ash in a molten, slag form. One of the big advantages of entrained bed gasifiers is that they can utilize any type of coal. As shown in Table 4, Koppers-Totzek, Texaco and Shell-Koppers are all entrained-bed gasifiers.

In the Koppers-Totzek gasifier pulverized coal is horizontally injected with steam and oxygen into the reactor which is essentially operating at atmospheric pressure. The gasification temperature is around 2700°F. At this high temperature, the ash is in a molten slag form which drops into a quench tank and is removed.[1]

The Koppers-Totzek gasifier is a first generation technology which, like the Winkler and Lurgi, has had extensive commercial experience and therefore is considered proven and available technology.[20,14,15]. Five of the 24 proposed projects submitted to the Synthetic Fuels Corporation plan to use Koppers-Totzek gasifiers which would seem to confirm its reliability. It will handle all types of coal but does require large raw gas compressors since it operates at atmospheric pressure.

The Texaco gasifier is a coal-slurry fed, high-capacity gasifier which handles all types of coals and produces very little by-product. The slurry which is composed of pulverized coal and water is pumped with oxygen into the top of the high-pressure (600-700 psig) gasifier and fired downwards. The product gas is withdrawn through a side nozzle at a temperature around 2500°F. The molten slag is removed through a slag hopper beneath the quench chamber.[1]

Since the coal is fed in a water slurry the coal does not have to be dried. This can be a big advantage over gasifiers (predominantly for western coals) which use part of their coal to dry the rest of the feedstock. Drying is expensive, it reduces efficiency and it raises operating costs.

The high operating pressure is also an advantage since the synthesis gas must be fed at even higher pressures to the methanol

unit. Although the operating cost for high pressure may be higher, it more than makes up for the high cost of compressors needed with low pressure gasifiers.

However, in order to have good efficiency the solid content of the slurry feed must be high, 50-60 percent. When lignite is slurried with water, the highest untreated solid concentration is about 43 percent because lignite naturally contains up to 35 percent moisture. If the lignite is pretreated, the moisture content can be lowered to more efficient levels.[9] The drawback is that pretreatment is an added cost to production, (although not too large).

Another disadvantage of the Texaco process is that it requires more oxygen than most of the other processes.

Although the Texaco gasifier has not yet been used on a commercial scale it has been extensively tested at a pilot plant in Montebello, California[1] and at three demonstration plants: the Ruhrchemie/Ruhrkohle plant in Oberhausen, West Germany; 2) Tennessee Valley Authority's ammonia-from-coal plant in Muscle Shoals, Alabama; and 3) an air blown gasification plant at a chemical facility in the USA.[17] Texaco appears to be the leading second generation technology and being planned for two projects already underway: Tennessee Eastman's project in Kingsport, Tennessee to produce acetic anhydride and other chemicals from methanol made from coal, and Southern California's Cool-Water power generation station in Daggett, California.[15]

The Shell-Koppers gasifier is very similar to the Texaco gasifier in that it can also use any coal and produces very little byproduct. However, it is likely that the process will not be commercialized for a couple of years since only limited data is available on a 150 ton per day demonstration plant.[14]

One of the gasifiers that was used in the design studies to be reviewed later was a Foster-Wheeler entrained bed gasifier. This gasifier unit consists of two stages. In the second which is an entrained gasifier operating at 300-400 psig and 1700°F, transport gas from stage one and pulverized raw coal are introduced yielding slag and the product gas. The char which is removed from the product gas is then sent with steam and oxygen to the first stage producing the transport gas which is recycled to the second stage.[1] As of 1977 the Foster-Wheeler gasifier was in the early stages of pilot plant development.[1]

The gasifier used by Badger was a version of an entrained bed gasifier.[4] According to Badger this gasifier is operated in an oxygen-blown mode with a molten slag-bath at the bottom. The gasifier has a total of 14 feed nozzles; 6 for coal and lime, 6 for oxygen and steam, and 2 for recycled char. The nozzles, which are distributed around the periphery of the vessel, fire tangen-

tially and at a 45 degree angle toward the surface of the slag to make it rotate. Dense-phase pulverized coal and lime are pneumatically fed into the lower section of the gasifier which operates at 500 psig. The lime is a fluxing agent which is added to obtain a slag viscosity of 10 poise. The oxygen and superheated steam are added as gasifying agents. The coal, which is partially pyrolyzed in the reaction, is gasified at 3000°F.

The advantages that are claimed in the literature for this gasifier are that it can handle any type of coal and that the raw gas is free of tar and high boiling hydrocarbons. When Badger compared dry and wet (slurry) feeding they found that dry feeding was economically superior to the slurry feed because the slurry feed required 29 percent higher coal feed and a 73 percent higher oxygen feed for a given synthesis gas, and therefore methanol, rate. Badger also found that a steam-oxygen gasification medium produces the highest thermal efficiencies ([4] pg. 64,65).

According to Badger "this single shaft high pressure slag-bath gasifier is based on published information for entrainment and other types of gasifiers and for the Rummel/Otto Gasifier which is proven at atmospheric pressure. It is a new concept and further development work may be necessary. Similar gasification principles have been studied and pilot plant tests have been conducted at lower pressures. Mechanical problems are recognized and are believed to be solvable".[4]

Until recently, industry has been very sluggish in its progress to reimplement coal gasifiers in the U.S. However, the increasing cost of natural gas has sparked a new interest in coal gasification and the majority of the coal or shale-based synthetic fuel projects currently being planned use coal gasification.[31] Table 6 lists some of the current projects which are now planned or proposed.

One example is the previously-mentioned Cool Water combined-cycle power-generation demonstration plant, to be located in Daggett, California. It will gasify 1000 tons per day of coal to produce 100 MW of electricity. According to a recent article in Chemical Engineering Progress, the facility, which will use the "proven" Texaco Coal Gasification Process, is to begin construction in July 1981 and be ready for start-up before the end of 1983.[32]

V. Synthesis Technology

The purpose of this section is to review available indirect liquefaction processes with the emphasis being placed on commercial feasibility, process description (reactor configuration, operating conditions, etc.), product quality, and a comparison of technological advantages and limitations. The processes that have been reviewed include seven methanol synthesis technologies (ICI,

Table 6

Coal to Methanol Projects

<u>Project Name</u>	<u>Plant Size (Barrels Methanol/day)</u>	<u>Construction Date</u>	<u>On Stream Date</u>
1. Great Plains Coal Gasification Project Mercer County, ND	125	July 1980	1984
2. Coal-to-Methanol-to Acetic Anhydride Tennessee Eastman Kingsport, TN	4,200	1980	1983
3. *Beluga Methanol Project, Granite Point, AK	54,000	1982-1983	1987
4. Grants Project **(ETCO), Grants, NM	3,608	1982	1984-1985
5. Mapco Synfuels Carmi, IL	35,000	1982	1987
6. Peat-to-Methanol **(ETCO), Creswell, NC	3,714	1982	1984
7. Keystone Project Cambria and Somer- set Counties, PA	100,000	1983	1986
8. Dunn Nokota Lignite-to-Methanol Dunn County, ND	85,000	1983	1987
9. Chokecherry **(ETCO), Moffat County, CO	3,608	1982	1984-1985

* Feedstock is 60 percent natural gas, 40 percent coal

** Energy Transition Corporation (ETCO)

Source: EPA [8]

Lurgi, Haldor Topsoe, Mitsubishi Gas Chemical, Vulcan-Cincinnati, Wentworth Brothers' and Chem Systems) and two gasoline/petroleum synthesis technologies (Mobils' Methanol-to-Gasoline and Fischer-Tropsch).

The results of this section are briefly summarized in Table 7. Of the seven methanol synthesis processes that were examined, the ICI, Lurgi, Haldor Topsoe, Mitsubishi Gas Chemical and Vulcan-Cincinnati technologies have several commercial scale processes in operation today. The Wentworth Brothers' methyl fuel process is adapted from proven technologies and may be close to commercialization.[13] The Chem Systems process is not commercially feasible at this time since it is only at the pilot plant stage. The latest report on the Mobil MTG process[19, 3/80] was that the 4 barrel per day (BPD) pilot plant was the biggest operating unit to date, but that plans for 100 BPD and 13,000 BPD plants were proceeding. Mobil states that MTG is ready for commercialization[19], but at this time the MTG process is not commercially proven. The Fischer-Tropsch process, which has been operating for 25 years in South Africa, is unquestionably proven.

From a process point of view the high pressure methanol synthesis technologies (Vulcan-Cincinnati and Wentworth Bros.) are better suited for large scale production plants whereas the low pressure methanol synthesis technologies (ICI, Lurgi, Haldor Topsoe, Mitsubishi and Chem Systems which operate between 30 and 130 atm) can be used with any size plant.[13] This is because the high pressure plants have high throughputs which tend to compensate for the higher cost for compression, especially for very large plants. Although individual efficiencies have not been reported for the methanol synthesis processes it is probable that many of the technologies have comparable efficiencies since they are highly developed and very competitive. Two big factors that affect efficiency are the extent of heat recovery and the percent conversion of carbon monoxide and hydrogen to methanol per pass in the converter. Of the methanol technologies listed most have a conversion per pass of about 5 percent (e.g., ICI) while the Chem Systems process claims up to 20 percent.[13] Concerning heat recovery Lurgi claims to be more efficient than ICI because it uses a heat exchanger type reactor versus the quench type used by ICI.[1] Since Chem Systems uses a liquid phase process it should get an even higher recovery of heat than Lurgi. Over the past ten years ICI and Lurgi (more recently) have dominated the methanol synthesis market.[1,25] Since these two processes are so competitive it would seem logical that their economics would be the same, and compared to other commercial processes, be comparable if not less expensive.

Parsons has stated that the Chem Systems process shows a slightly higher thermal efficiency and slightly lower capital cost than Lurgi and ICI synthesis; however, they believe room for improvement over these synthesis units is small.[1] A comparison

Table 7

Methanol Synthesis Processes

<u>Vendor</u>	<u>Catalyst</u>	<u>Pressure (atm)</u>	<u>Temperature (°C)</u>	<u>Reactor Type</u>	<u>Cooling</u>
ICI*	Cu/Zn/Al	50-100	220-290	Single fixed-bed	Multiple gas quench
Lurgi*	Supported Cu	30-50	235-280	Tube in shell	Steam generation
Topsoe*	Cu/Zn/Cr	50-100	220-350	Radial flow	Boiler-feed-water heating
Vulcan-Cincinnati	Zn/Cr	300-350	300-400	Multiple bed	Cold-shot quench, plus external gas cooling
Mitsubishi Gas Chemical*	Cu/Zn/Cr	50-130	240-310	-	Multiple gas quench
Wentworth Bros.*	Multi-Catalyst	up to 400	200-400	-	-
Chem Systems	Cu/Zn	34-68	215-250	Liquid entrained and liquid fluidized	Recirculated inert hydrocarbon liquid
Mobil MTG	Zeolite	2.7	330-400	Fixed or fluid	
Fischer-Tropsch*	Cobalt or Iron	0-25	200-325	Fixed and fluid with cooling tubes	Steam generation

* Proven on a commercial scale.

of the Wentworth Brothers' process (using available published information) with other processes did not show any inherent economic advantage for the WBI process.[13]

A. ICI Low-Pressure Methanol Synthesis

Status: The ICI low pressure (50-100 atm) methanol synthesis process is commercially proven worldwide.[6,13,22,23,24,33]

Process: Feed gas consisting of an approximately 2 to 1 ratio of hydrogen to carbon monoxide is fed into the synthesis loop. The methanol conversion is highest when the hydrogen to carbon monoxide ratio is 2 to 1 and the carbon monoxide to carbon dioxide ratio is as high as possible.

The first part of the synthesis consists of desulfurizing the feed gas when necessary to prevent the highly sensitive copper-based catalyst from being poisoned. This is accomplished by passing the gas through sulfur guard beds, which are typically made of zinc oxide (or, less commonly, activated carbon) to achieve sulfur levels below 1 ppmv.

The feed gas is then compressed to the recycle loop pressure, mixed with the recycle gas and then compressed to reactor pressure as it enters the methanol converter. The converter is a pressure vessel containing a bed of catalyst.

The temperature of the bed is controlled by the extent of the exothermic reaction and the quenching of the reaction by cold feed gas. The pressure range is 50 to 100 atm while the temperature must be kept below 300°C (210 to 300°C) since the catalyst becomes deactivated at higher temperatures. The exit gas is passed through heat recovery units for initial cooling and then sent to the methanol separation unit where a crude methanol product is produced (95 percent methanol by weight). Conversion of CO to methanol per pass is about 5 percent.[13]

Catalyst life at pressures of 50 to 60 atm is 4 years while maximum catalyst life at 100 atm is 2.5 years (average is 1-2 years).

Advantages/Disadvantages: Compared with high-pressure processes, the ICI process is more adaptable to both large and small plants (55 to 2750 TPD) whereas high pressure processes are limited to large plants (1,000 to 2,500 TPD). Compression costs are lower because of reduced pressure.

The disadvantages are that the high-pressure processes allow a higher throughput of gas for the same size reactor and that the catalyst cost for low pressure technology is five times as expensive (50 cents/ton vs. 10 cents/ton).[13]

B. Lurgi Low-Pressure Methanol Synthesis

Status: The Lurgi low-pressure synthesis process (30-50 atm) is commercially proven.[13,25,23,22,33]

Process: The Lurgi methanol synthesis process uses a shell-and-tube reactor. The copper-zinc catalyst is packed in vertical tubes contained within a reactor shell which is filled with boiling water. The exothermic heat of reaction is removed by the generation of steam, thereby controlling the temperature of the reactor.

The hydrogen to carbon monoxide ratio of the feed gas is normally between 2 and 3, whereas the ratio of (hydrogen minus carbon dioxide) to (carbon monoxide plus carbon dioxide) is held around 2.2. After desulfurization the feed gas is compressed, combined with recycle gas and preheated before being fed into the reactor at one specific location. The Lurgi reactor has an operating range of 30 to 100 atm and 200 to 300°C but is typically operated at 70 atm and 260 to 270°C. The exit gas contains about 4-6 percent methanol and is sent to condensers to recover the crude methanol product which is generally sent on for purification.[13]

Advantages/Disadvantages: Like other low-pressure processes the Lurgi process has an economic advantage over high-pressure processes due to decreased compression costs at lower pressure. The reactor design also permits direct recovery of the exothermic heat of reaction by steam generation rather than a partial quench of the reaction to control heat build up.[13]

According to Lurgi, a natural gas to methanol plant using their synthesis technology is more efficient[22] and consumes 3-5 percent less natural gas per ton of pure methanol than competing technology. They estimate the annual savings for a 2,500 stpd plant is \$4.2 - 5.0 million (U.S. dollars).[25]

When Badger compared the low pressure (1500 and 1400 psig) methanol synthesis processes employing quench type converters and licensed by Imperial Chemical Industries and Mitsubishi Gas Chemicals Corporation with that of Lurgi's tubular type low pressure (750 and 1200 psig) process they selected Lurgi's process for two reasons:

- Lower investment and operating costs for syn-gas compression
- Maximizes medium pressure steam production; thus minimizing overall utility costs and pay out time.

C. Haldor Topsoe Methanol Synthesis

Status: The Haldor Topsoe Methanol Synthesis process is commercially proven.[13,22,33]

Process: The Haldor Topsoe process is similar to other low- and intermediate-pressure methanol synthesis processes. The synthesis utilizes a copper-zinc-chromium catalyst in two or three radial flow converters operated in series. After being desulfurized (20 ppbv sulfur level) with zinc oxide guard beds, the feed gas is mixed with recycle gas and passed through the reactors flowing radially outward through each catalyst bed. Operating pressure and temperature ranges are respectively 48 to 144 atm and 220 to 350°C. The exothermic heat from each reactor is recovered by heating boiler feedwater with the hot exit gases. The gases are then condensed and sent to a separator where crude methanol is separated from uncondensed gases and later sent to product upgrading.[13]

Advantages/Disadvantages: The Haldor Topsoe process can operate at intermediate pressures for higher throughputs. It can also operate at higher temperatures which increases the activity of the catalyst provided it can retain its active sites and structural integrity.

D. Mitsubishi Gas Chemical Methanol Synthesis

Status: The Mitsubishi Gas Chemical (MGC) methanol synthesis process is commercially proven.[13,33]

Process: The MGC process appears to be very similar to ICI's intermediate-pressure process since both designs use a quench converter with a ternary copper-based catalyst operated at low temperature and intermediate pressure (240-310°C and 50-130 atm). The feed gas is split into a feed stream which is heated and fed into the converter, and a quench stream which is injected at several bed levels to control the buildup of the exothermic heat of reaction. After being used to preheat the feed gas the exit gas is condensed and sent on to distillation for a product purity in excess of 99 percent methanol by weight. Part of the recycle gas is used for fuel. The catalyst has an expected life of just over 1 year since it is very sensitive to sulfur.

Advantages/Disadvantages: The MGC intermediate pressure process has the advantage of accomodating moderately higher throughputs than lower pressure processes while keeping compression costs down as compared to high pressure processes.

The ICI catalyst appears to have a longer catalyst life (2 years for ICI vs. greater than 1 year for MGC). The MGC process typically uses a higher hydrogen to carbon oxides ratio in the feed gas than other processes (3.1 compared with 2.2 for other processes) but this is because it has only been used when natural gas is the feedstock.[13]

E. Vulcan-Cincinnati High Pressure Methanol Synthesis

Status: The Vulcan-Cincinnati high-pressure process is commercially proven.[13,33,22] However, the company had to stop operation in 1973 when the Middle East war forced the cancellation of a very large methanol plant in Saudi Arabi in which Vulcan had heavily invested (see Wentworth Brothers' process below).

Process: The feed gas ratio for $H_2/(CO + 1.5 CO_2)$ should be adjusted to a value of 2 for optimum conversion after desulfurization. The feed gas is then compressed and fed to the converter which is usually operated in the range of 340 to 400°C and 200 - 300 atm. The convertor operates adiabatically with considerable temperature rise due to the exothermic heat of reaction, which is controlled by quenching the reaction with cold feed gas at several levels. After conversion the crude methanol product is condensed for removal yielding a product containing up to 97 wt percent methanol (depending on feed gas composition). There is also an option of producing up to 20 wt. percent of higher alcohols by changing operating conditions which would be helpful if used for blending with gasoline.

The catalyst, which is poisoned by H_2S levels greater than 3 to 5 ppm, has a typical life of about 4 years and can be regenerated. Conversion of CO to methanol per pass is approximately 5 percent.

Advantages/Disadvantages: The high pressure process is well suited for large methanol synthesis trains due to the high throughputs occurring. Catalyst costs are also less than the low-pressure copper-based catalysts, are not as sensitive to sulfur and can be regenerated. The process can produce a wider range of fuel products (3 - 20 percent higher alcohols).

Some disadvantages are: 1) that the high-pressure process has greater compression costs, 2) that the catalyst requires higher temperatures and pressures because it is not as active as the copper-based catalysts, and 3) that it may not be suited for small methanol plants.[13]

F. Wentworth Brothers Methyl Fuel Process

Status: The term methyl fuel, copyrighted by Vulcan-Cincinnati, represents the product of a methanol synthesis process which is focused on producing methanol for fuel rather than chemical uses. After Vulcan-Cincinnati stopped operation in 1973, the Wentworth Brothers and other engineers from Vulcan formed a new corporation in May 1975. Based on Vulcan experience and technology and relying on catalyst improvements and a reactor design adapted from proven petroleum technology, Wentworth Brothers, Inc. (WBI), is now marketing what they believe to be a much improved process for the production of large quantities of fuel-grade

methanol. No commercial plants are in operation, but short-term tests of the catalyst at 300 TPD in a commercial methanol train have reportedly verified the basic operating parameters for the WBI methyl fuel process.[13]

Process: Details of the process operation, catalyst formulation, and reactor configuration are considered proprietary and are not available.

Advantages/Disadvantages: What is known is that the new catalyst is reportedly more active and durable than conventional low pressure catalysts although at the expense of the selectivity for methanol. The more active catalyst allows operation of the process at increased space velocities (throughput per reactor volume), and at higher temperatures and pressures which maximize fuel production per reactor train. The range of operating conditions for the methyl fuel process includes pressures up to 4,000 psi (270 atm) and temperatures from 200 to 400°C. The catalyst is claimed to be effective at CO₂ concentrations ranging from 20 percent to essentially zero versus conventional copper-based catalysts which require some CO₂. Oak Ridge National Laboratory states that from available published information it is not possible to ascertain whether the WBI Methyl Fuel Process has any inherent economic advantage over conventional copper-based methanol synthesis processes.[13]

G. Chem Systems Synthesis

Status: As of August 1980, the Chem Systems process was ready to move to the pilot plant stage.[21]

Process: The major difference between the Chem Systems synthesis and the other synthesis processes is that an inert hydrocarbon liquid is used as the medium for the catalyst instead of a gaseous phase. This liquid phase allows high conversions of carbon monoxide and hydrogen to methanol in addition to maximum recovery of reaction heat.[21]

In the process synthesis gas containing carbon monoxide, carbon dioxide and hydrogen is passed upward into the reactor concurrent with the inert hydrocarbon liquid, which is recovered in the separation plant and recycled back to reactor with the unconverted synthesis gas.[1,21] The fuel grade methanol product is 95-96 percent methanol by wt.[13]

Advantages/Disadvantages: Chem Systems claims that their conversion to methanol per pass is about 4 times as great as other processes (20 percent vs. 5 percent). However, Parsons believes that this process only has slight cost advantages over the existing processes.[1][13]

Ralph M. Parsons states that a new catalyst formulation with superior mechanical strength still needs to be developed to make the liquid phase methanol synthesis viable.[1] Breakdown of catalyst, inhibition of catalyst by fluid and insufficient solubility of the synthesis gas in the fluid are other possible problem areas with this design.[22]

H. Mobil Methanol-To-Gasoline (MTG) Process

Status: Mobil has conducted developmental studies of this process in fixed- and fluid-bed bench-scale units with two reactors being used in the fixed bed unit. The fixed bed reactor achieved over 200 days of successful operation. The single reactor fluid-bed unit has undergone two months of testing. Since the fluid bed reactor had a number of advantages over the fixed bed reactor, a 4-BPD fluid-bed pilot plant was designed, built and operated under a follow-on DOE contract in 1976-78. Startup and operation of this fluid-bed unit were reported to be very successful. Plans are currently under way for a 100 BPD fluid bed pilot plant sponsored by DOE, the Federal Republic of Germany, German industrial participants, and Mobil.[13,18] Of interest also is the reported news that, since November 1979, the government of New Zealand has been pursuing the Mobil methanol-to-gasoline process, with negotiations proceeding for a 13,000 BPD fixed-bed unit for installation almost immediately.[18,19]

Process: The conversion of methanol to hydrocarbons and water is a very exothermic reaction giving off 740 Btu/lb of methanol. Heat removal is therefore the principal problem in designing a reactor system.[7] For the fixed bed reactor the problem is minimized by dividing the reaction into two steps and using two reactors in series. In the first reactor crude methanol is partially dehydrated to an equilibrium mixture of methanol, dimethyl ether and water over a dehydration catalyst[14,19] releasing about 20 percent of the reaction heat.[7] In the second reactor the new shape-selective zeolite catalyst is used to convert both methanol and dimethyl ether to a liquid hydrocarbon product. This hydrocarbon liquid product is then sent to a fractionation unit where a deethanizer sends the ethane rich overhead product to the SNG train and the bottoms are sent to a stabilizer. The overhead product of the stabilizer is composed of isobutane and butene/propene which are sent to an alkylator to produce more gasoline and commercial grade propane and butane. The bottoms product of the stabilizer is a stabilized gasoline which is mixed with the gasoline product from the alkylation unit and sent to the gasoline blending unit to yield a high octane (93 research octane) gasoline.[7]

The inlet temperature of the second reactor is about 625°F. The adiabatic fixed bed process operates at essentially 100 percent conversion of methanol to hydrocarbons and water until the catalyst deactivates by carbon formation to an activity level

where only partial conversion of methanol is achieved.[19] The zeolite catalyst must be regenerated once every 14 days.[7]

In the fluid bed process one reactor is used, operating at 750°F and 40 psig. The hydrocarbon product is generally treated in the same manner as the fixed bed product with the exception of a few changes. A deethanizer absorber is used in place of the high pressure deethanizer tower to provide a recycle stream to the reaction for increased propane-plus yield. A rich oil tower is also required.[7]

Advantages/Disadvantages: For the fluid-bed reactor the methanol conversion is greater than 95 percent, producing about 44 percent hydrocarbons and 56 percent water. The pentane-plus gasoline fraction of the hydrocarbons is about 60 percent. The propene, butene, and isobutane produced are approximately the right proportions for alkylation, bringing the total yield of 9 lb Reid Vapor Pressure gasoline (96 unleaded RON) up to 88 percent of the total hydrocarbon yield. The thermal efficiency for the methanol conversion is quoted at 95 percent.[13,27]

One potential problem with the gasoline produced from both of these processes is the presence of durene which boils in the gasoline range but has a freezing point of 175°F. This could cause engine problems since the durene could crystallize out in an engine's carburetor. Durene is present in conventional gasoline in very small amounts but could be present in relatively large amounts (3-6 percent) in Mobil MTG-gasoline. Durene levels of 5 percent in gasoline did cause some unsatisfactory engine operations during tests but at 4 percent levels effects were minimal. Since durene levels can be maintained to acceptable levels by proper process controls and it could always be mixed with conventional gasoline[28] the presence of durene may not pose too much of a problem.

I. Fischer-Tropsch

Status: The Fischer-Tropsch process is proven technology which has been producing liquid hydrocarbons at SASOL in Sasolberg, South Africa since 1955.[7] This is the only commercial scale coal liquefaction plant operating in the world today. The current SASOL plant uses two reactor schemes, a fixed-bed and a fluid bed. The current SASOL expansion to 50,000 BPD is based on a fluid-bed design.

Process: With the fluid bed reactor, purified synthesis gas is compressed and charged into the reactor. After mixing with the circulating hot iron catalyst, the reaction takes place as the mixture flows up the reactor through tube bundles in which oil is pumped for heat removal. At the top of the reactor, the mixture enters a large vessel in which cyclones separate the iron and vapor. The hot oil is circulated to a steam generator where 200

psig steam is produced. The overhead vapor is condensed and the vapor split into a recycle and purge stream, the latter being sent to hydrocarbon recovery. The condensed liquid is split into a cold recycle liquid and a light oil product. Based on a 2.13 molar feed ratio of $H_2:CO$ to the F-T synthesis units the final product yield consists of 24 percent SNG, 54 percent 10 RVP gasoline, 9.8 percent diesel fuel, 5.5 percent alcohols, 3.8 percent LPG and 2.8 percent heavy fuel oil (percentages on a HHV BTU basis).[7]

The commercial catalysts include cobalt, for the fixed-bed reactor, and iron for both the fixed- and fluid-bed reactors. Operating conditions range from 200-325°C and from atmospheric pressure to 25 atm depending on the desired products. Because all of the reactions in the F-T process are exothermic, heat removal is important. For the fixed bed reactor, designs include a heat exchanger being cooled by boiling water or circulating oil. Fluid bed reactors use internal tube bundles for reaction heat removal.[7]

Advantages/Disadvantages: According to a study done by Mobil[7] which compares the F-T process to the Mobil MTG process, the MTG process has a number of advantages over F-T. For example, the Mobil technology gives a higher liquid product/SNG ratio (energy basis), 47/53 vs. 35/65. The F-T route has 18 processing steps compared with nine for MTG. The MTG gasoline had a higher RON (93 vs. 91) and a lower olefin content (11 vs. 20 percent). The economics show that the MTG gasoline cost is moderately cheaper than F-T gasoline (\$.60 - \$1.00 per gal. vs. \$.70 - \$1.35 per gal.). The overall efficiency for MTG was 62 percent vs. 58 percent for F-T.[7]

A recent article in Hydrocarbon Processing advocating F-T stated that the 40-45 percent efficiency reported for the operating F-T process is really depressed because the SNG produced is reformed to make additional H_2 and CO instead of being sold as a product (SNG is not marketable in South Africa). If the process was brought to the U.S. the author believes that SNG would be a viable product as a pipeline quality methane gas, thus increasing the F-T efficiency to 60 percent (close to 58 percent which was previously stated by Mobil). In addition the article points out that coal consumption for the F-T process is high only because they use a low quality coal with 30 percent ash.[26]

VII. Comparison of Indirect Liquefaction Design Studies

As discussed earlier, there have been a number of studies evaluating the indirect liquefaction of coal. This chapter is divided into four main sections, the production of methanol from: 1) bituminous coals, 2), subbituminous coals, and 3) lignite and 4) the production of gasoline from methanol using the Mobil MTG technology; however, this fourth section also includes the production of gasoline and other products via Fischer-Tropsch technology.

Each section begins with a brief introduction and a comparison of three aspects of the various studies: level of engineering design, feedstock analysis, and material balances and efficiencies. Then capital, operating and product costs for each study will be presented on a consistent economic basis and then compared to reconcile as many of the differences as possible.

A. Methanol from Bituminous Coal

There were five original studies available which investigated the technical feasibility of producing methanol from bituminous coals:

1. R.M. Parsons, Co., for EPRI, "Screening Evaluations: Synthetic Liquid Fuel Manufacture,"[1]
2. C.F. Braun for EPRI, "Coal to Methanol Via New Processes Under Development: An Engineering and Economic Evaluation,"[2]
3. Dupont Co. for U.S. ERDA, "Economic Feasibility Study, Fuel Grade Methanol from Coal for Office of Commercialization of ERDA,"[3]
4. Badger Plants, Inc., "Conceptual Design of a Coal-to-Methanol Commercial Plant,"[4] and,
5. Exxon Research and Engineering Co., "Production Economics for Hydrogen, Ammonia, and Methanol During the 1980-2000 Period."[5]

The studies differ in depth of design and use different assumptions with respect to key economic parameters; also plant sizes vary widely. Since methanol-from-coal technology is well developed and much of it is common to all of the studies, the large cost differences between the studies should be reconcilable by placing them on a consistent economic basis.

Depth of Design: The level of engineering detail of the five studies varies. The most detailed studies were found to be those performed by Badger and EPRI/C. F. Braun. These studies include complete material balance information for a detailed flowsheet and extensive design details, including drawings. The studies with the next highest level of engineering design are the EPRI/Ralph M. Parsons and Du Pont studies. These studies are screening evaluations and their level of detail is not sufficient to allow comparison with more detailed studies. The EPRI/Ralph M. Parsons study includes evaluations of four gasification processes in combination with Chem Systems Methanol synthesis; these four processes are denoted as Cases 1, 2, 3 and 4 of the Parsons study. The study found to be based on the least detail is the Exxon/Chem Systems study which consists of summary economic information for

two gasification processes, but provides no details for material and energy balances. These two processes are denoted as Cases 1 and 2 of the Exxon study.

Ultimate Analysis for Bituminous Coals: Ultimate analyses of the bituminous coals used in the various studies are presented in Table 8. Also listed in this table are the heating values of the coals. The Exxon/Chem Systems study did not report an analysis for their coal; however, they used an Illinois high sulfur coal which is probably similar to the Illinois No. 6 bituminous coals reported in the two EPRI studies. The Du Pont coal is also similar to the coals presented in the EPRI studies. However, the coal considered in the Badger report is a low sulfur coal which would meet the sulfur dioxide standard for large power plants.[12] It is unlikely that a coal of this quality would be used to produce methanol.

Material Balance and Efficiencies: Feedstock and product rates for each study are presented in Table 9. Methanol and coal are presented on both a short ton per calendar day (tpd) basis and on an energy basis. Other products include fuel gas in Cases 1 and 2 of the Parsons study; the Badger study includes chemical grade methanol along with the fuel grade. All rates are based on 50,000 FOEB/CD of total products.

Table 9 shows that the process efficiencies for the studies vary from 49.3 percent for the Koppers-Totzek/ICI case prepared by Exxon/Chem Systems to 58.2 percent for the Texaco/Chem System case prepared by EPRI/Parsons. An investigation of these efficiencies shows that the two Koppers-Totzek cases are amongst those with the lowest efficiencies. This is not surprising since the Koppers-Totzek gasifier is the only first generation gasifier listed in this table, and since the gasifier operates at near atmospheric pressure so that the product synthesis gas must be compressed, thus resulting in an efficiency penalty. When comparing the three Texaco gasifier studies it can be seen that there is a substantial difference (7.5 percent) between the highest and lowest efficiency. The C.F. Braun study reports an efficiency of 55.7 percent whereas the DuPont study reports an efficiency of 50.6 percent. A key variable in the Texaco process which affects the efficiency is the coal/water slurry concentration. The greater the coal concentration in the slurry, the higher the efficiency of the Texaco process.[9] The C.F. Braun case utilized a 59 percent coal concentration in the slurry whereas the DuPont study utilized a 54 percent coal concentration. This difference will account for part of the efficiency discrepancy. The Parsons/EPRI study did not report any processing information for the Texaco gasifier and therefore the high efficiency reported for this process could not be investigated. The remainder of the processes reported in Table 9 utilize other advanced technology gasifiers, and therefore are expected to have higher efficiencies than first generation gasifiers, e.g., Koppers-Totzek.

Table 8

Ultimate Analysis of Bituminous Coal Feedstocks

<u>Study:</u>	<u>Parsons</u> <u>Cases 1,2,3,4[1]</u>	<u>C. F. Braun[2]</u>	<u>DuPont[3]</u>	<u>Badger[4]</u>	<u>Exxon[5]</u> <u>Case 1 and 2</u>
<u>Coal Type</u>	<u>Bituminous</u> <u>Ill. No. 6</u>	<u>Bituminous</u> <u>Ill. No. 6</u>	<u>Bituminous</u>	<u>S. Appal.</u>	<u>Ill. High</u> <u>Sulfur</u>
HHV (Btu/lb.)	12,235 (wet)	12,150 (dry)	12,113 (dry)	12,840 (dry)	11,390 (wet)
LHV (Btu/lb.)	11,709	-	10,874	-	-
<u>Ultimate Analysis,</u> <u>Wt % Dry Coal</u>					
C	69.5	68.25	66.89	73.8	
H	5.3	5.00	4.47	4.8	
O	10.0	11.23	8.41	6.4	
N	1.3	0.81	1.28	1.6	
S	3.9	3.88	4.47	1.1	
Ash	10.0	10.83	14.48	12.3	
Total	100.0	100.0	100.0	100.0	
Wt% Moisture (As Recieved)	4.2	10	6.38	2.4	

Table 9

Methanol from Bituminous Coal:
Feedstock and Product Rates
(Normalized to 50,000 FOEB/CD Product)

Study:	Parsons[1]		C.F. Braun[2]		DuPont[3]		Badger[4]		Exxon[5]	
	Case 1	Case 2	Case 3	Case 4	Case 1	Case 2	Case 1	Case 2	Case 1	Case 2
<u>Mass Basis</u>										
<u>Feedstocks</u>										
Coal, tpd	21,722	21,150	23,006	20,714	21,795	25,685	20,048	24,447	26,188	
<u>Products</u>										
Methanol, tpd	14,500	15,132	15,349	15,349	15,172	16,223	14,570	15,227	15,227	
HHV Btu/lb	9,610	9,610	9,610	9,610	9,722	9,092	9,407	9,687	9,687	
Fuel Gas, mscf/CD	20.39	4.89	-	-	-	-	-	-	-	
Chemical Grade Methanol, tpd	-	-	-	-	-	-	1075	-	-	
<u>Energy Basis (HHV), mBtu/CD</u>										
<u>Feedstocks</u>										
Coal	531,531	517,544	562,977	506,873	529,623	582,543	514,834	556,902	596,565	
Electricity (energy equivalent)	-	-	-	-	-	-	-	1810	1812	
<u>Products</u>										
Methanol (Fuel Grade)	278,682	290,845	295,000	295,000	295,000	295,000	274,121	295,000	295,000	
Fuel Gas	16,317	4,155	-	-	-	-	-	-	-	
Methanol (Chemical Grade)	-	-	-	-	-	-	20,879	-	-	
Thermal Eff'cy, %	55.5	57	52.4*	58.2	55.7	50.64	57.3	52.8	49.3	

* 95% conversion assumed for gasifier as opposed to 100 percent for the other cases.

Economics: In this sub-section the capital and operating costs for each study will be presented in order to obtain the desired product costs. These costs have been placed on a consistent economic basis as reported in a previous report.[10a]

Table 10 presents all of the investment costs (normalized to 50,000 FOEB/CD) broken down into individual process unit costs. This table shows that the total instantaneous investments range from \$1.62 billion for the Shell-Koppers/ICI case of the Exxon study to \$2.56 billion for the Koppers-Totzek case of the Parsons study, which represents a \$940 million instantaneous investment difference.

Operating costs are presented in Table 11. Some of the studies did not itemize the operating costs which makes it difficult to compare these costs between each study. The net annual operating costs range from about \$315 - 480 million. Most of the operating cost estimates lie in the \$340 - 440 million per year range.

Tables 12 and 13 present economic summaries of methanol costs for capital charge rates of 11.5 and 30 percent. For the lower capital charge rates product costs vary from \$5.30 for the Parsons/BGC Lurgi study to \$7.23/mBtu for the Parsons/Kopper-Totzek study. For the higher capital charge rate product costs vary from \$8.74 to \$12.42/mBtu.

Now some of the capital and operating cost differences will be reconciled. First, one would expect the Koppers-Totzek cases to have a high capital cost because it is a low pressure process; this results in higher compression and gasification costs than for high pressure gasification technologies. Also, since the Koppers-Totzek gasifier is a first generation gasifier, its capital cost is expected to be higher than the more advanced technology cases. The instantaneous plant investment for the Parsons/Koppers-Totzek study is \$2.56 billion while that of the Exxon study is \$1.9 billion, which represents about a \$650 billion investment difference. The investment estimate of the Exxon study is similar to those of the advanced technology gasification cases which seems unreasonable. The only difference between the two Koppers-Totzek cases is the methanol synthesis technology used. Since there is very little difference expected in the capital costs of the Chem Systems and ICI synthesis units, it is surprising to see such a large capital cost difference between these two cases, and the difference may be attributed to lack of design detail for the Exxon study which was discussed earlier. Therefore, it is believed that the Parsons study is more representative for the Kopper-Totzek case.

Three studies investigated the use of Texaco gasification technology: Parsons, C.F. Braun, and DuPont. The respective plant investment costs for these studies are \$2.05, \$1.75 and

Table 10

Methanol From Bituminous Coals: Capital Costs Summary*
(Millions of First Quarter 1981 Dollars)

Study:	Parsons[1]					C.F. Braun[2]	DuPont[3]	Badger[4]	Exxon[5]	
	Case 1	Case 2	Case 3	Case 4					Case 1	Case 2
Technology:	Foster Wheeler/ Chem. Systems	BCC Lurgi/ Chem. Systems	Koppers Totzek/ Chem. Systems	Texaco/ Chem. Systems		Texaco/ Chem. Systems	Texaco/ ICI	Slag Bath/ Lurgi	Shell- ICI	Kopper Totzek/ ICI
Gasification/ Synthesis										
Coal Preparation	83	40	87	73		110	112	37	-	-
Tar and Phenol	-	99	-	-		-	-	33	-	-
Recovery										
Gasification	186	145	566	385		167	205	187	-	-
Shift Conversion	42	56	51	51		97	75	66	-	-
Acid Gas Removal	281	241	266	283		187	175	193	-	-
Sulfur Removal	14	14	15	15		-	-	16	-	-
Synthesis Gas	71	76	125	-		-	-	33	-	-
Compression										
Methanol Synthesis	170	175	177	177		164	190	197	-	-
Cryogenic Recovery	29	30	-	-		-	-	16	-	-
Sulfur Recovery/	28	29	29	29		26	-	24	-	-
Methanol Drying										
Oxygen Production	348	263	374	430		300	373	313	-	-
Steam and Power	68	80	197	106		123	302	90	-	-
Generation										
Subtotal	1320	1248	1887	1549		1174	1432	1205	775	951
General Facilities	198	187	283	232		306	209	453	597	666
and Offsites										
Contingency	228	215	325	217		222	246	249	206	243
Contractor's Fee	44	42	63	52		43	48	42	40	47
Total Instantaneous	1790	1692	2558	2050		1745	1935	1949	1618	1907
Plant Investment										

* Investment costs are based on 50,000 FOEB/CD of product.

Table 11

Methanol From Bituminous Coals: Operating Cost Summary
(Millions of First Quarter 1981 Dollars Per Year)

Study:	Parsons[1]					Exxon[5]			
	Case 1	Case 2	Case 3	Case 4	C.F. Braun[2]	DuPont[3]	Badger[4]	Case 1	Case 2
Technology:	Foster Wheeler/	BGC Lurgi/	Koppers	Texaco/	Texaco/	Texaco/	Slag Bath/	Shell-	Kopper
Gasification/	Chem. Systems	Chem. Systems	Chem. Systems	Chem. Systems	Chem. Systems	ICI	Lurgi	Koppers/	Totzek/
Synthesis								ICI	ICI
Raw Materials:									
Ill. No. 6 Coal	218	212	231	208	219	258	199	245	263
Catalysts and	10.6	10.6	12.1	10.6	20.7	15	3.4	4.1	4.1
Chemicals									
Utilities:									
Power	-	-	-	-	-	-	23	6.8	6.8
Process Water	-	-	-	-	-	-	-	19.4	20.6
Stack Gas Clean-up	-	-	-	-	-	-	-	8.7	12.1
Labor and Related:									
Labor	-	-	-	-	36.3	12	29.9	1.3	1.3
Supervision	-	-	-	-	10.9	-	1.5	4.1	4.1
Plant Overhead	-	-	-	-	-	-	2.9	1.7	1.7
Capital Related:									
Maintenance	-	-	-	-	40.1	97	-	63.4	74.5
General Plant	-	-	-	-	23	-	-	41.2	48.4
Overhead	-	-	-	-	47.4	41	-	42.1	47.5
Insurance and	-	-	-	-	-	-	-	-	-
Property Tax	-	-	-	-	7.3	8.1	7.2	6.8	8.0
Interest on Working Capital	7.5	7.1	10.7	8.8	-	-	-	-	-
Other Operating Costs	124	120	181	149	1.2	17	52	-	-
Gross Annual	360	350	435	387	406	458	319	445	492
Operating Cost									
By-Product Credit	(14.7)	(13.9)	(12)	(13.1)	(13.2)	(19.5)	(3.8)	(15.0)	(16.0)
Net Annual	345	336	423	374	393	429	315	430	476
Operating Cost									

Table 12

Economic Summary of Methanol from Bituminous Coal, CCR = 11.5%
(Millions of First Quarter 1981 Dollars)

Study:	Parsons[1]			C.F. Braun[2]	DuPont[3]	Badger[4]	Exxon[5]	
	Case 1	Case 2	Case 3				Case 1	Case 2
Total Instantaneous Investment	1790	1692	2558	2050	1745	1935	1949	1618 1907
Total Adjusted Capital Investment	2030	1919	2901	2325	1979	2194	2210	1835 2163
Start-up Costs	126	119	179	147	122	136	137	114 134
Pre-paid Royalties	8	9	13	10	10	10	10	10 10
Total Capital Investment	2164	2047	3093	2482	2111	2340	2357	1959 2207
Working Capital	125	118	179	147	122	135	137	113 133
Total Capital Requirement	2289	2165	3106	2629	2233	2475	2494	2072 2440
Annual Capital Charge	249	235	356	302	243	269	287	225 265
Annual Operating Cost	345	336	423	374	393	429	315	430 476
Total Annual Charge	594	571	779	676	636	698	602	655 741
Product Cost								
\$/FOEB of Product	32.55	31.29	42.68	37.00	34.85	38.25	32.99	35.89 40.60
\$/mBtu of Product	5.52	5.30	7.23	6.28	5.90	6.48	5.59	6.08 6.88

Table 13

Economic Summary of Methanol from Bituminous Coal, CCR = 30%
(Millions of First Quarter 1981 Dollars)

Study:	Parsons[1]		C.F. Braun[2]		DuPont[3]		Badger[4]		Exxon[5]	
	Case 1	Case 2	Case 3	Case 4	Case 1	Case 2	Case 1	Case 2	Case 1	Case 2
Total Instantaneous Investment	1790	1692	2558	2050	1745	1935	1949	1618	1907	
Total Adjusted Capital Investment	1998	1888	2855	2287	1947	2159	2176	1806	2128	
Start-up Costs	126	119	179	147	122	136	137	114	134	
Pre-paid Royalties	8	9	13	10	10	10	10	10	10	
Total Capital Investment	2132	2016	3047	2444	2079	2305	2323	1930	2272	
Working Capital	125	118	179	147	122	135	135	113	133	
Total Capital Requirement	2257	2134	3226	2591	2201	2440	2460	2043	2405	
Annual Capital Charge	640	605	914	777	624	692	738	579	682	
Annual Operating Cost	345	336	423	374	393	429	315	430	476	
Total Annual Charge	985	941	1337	1151	1017	1121	1053	1009	1158	
Product Cost										
\$/FOEB of Product*	53.97	51.56	73.26	63.07	55.73	61.42	57.70	55.29	63.29	
\$/mBtu of Product	9.15	8.74	12.42	10.69	9.44	10.41	9.78	9.37	10.73	

* One FOEB = 5.9 mBtu

\$1.94 billion. The total gasification costs for the C.F. Braun and DuPont studies are almost identical (about \$565 million) whereas the total gasification cost from the Parsons study is about \$240 million higher. The main difference in the C.F. Braun and DuPont studies lie in differences in the cost of the non-gasification equipment (oxygen production, offsites, etc.). It is difficult to determine which of the studies is most representative; based on level of design, the C.F. Braun study should be considered most representative. However, to be somewhat conservative in comparing methanol costs to direct liquefaction cost both the C.F. Braun and the DuPont studies will be chosen and a range of costs used. The product costs for these Texaco studies lie in about the same range as those for the other advanced technology cases, which seems reasonable.

In an effort to determine the representativeness of the Badger costs, DOE commissioned Oak Ridge National Laboratory in June, 1978 to make an independent assessment of the Badger report.[35] ORNL reported that Badger's design is based on equipment sizes well beyond the present state-of-the-art in order to take advantage of the projected economies of scale. Therefore, ORNL believes that the Badger design is more representative of an Nth plant design rather than a first plant design. For a first plant design the Badger capital and operating costs appeared to be unreasonable to ORNL. The operating cost for the Badger study listed in Table 11 are lower than any of the others listed in the table and does appear to be low. However, the Badger study is based on an advanced technology "slag bath" gasifier, and the capital cost based on this design is a bit higher than those of the other studies based on advanced technology gasifiers listed in Table 10. It is expected that these designs would have a lower capital cost than those based on technology commercially available or near commercially available. The degree to which these plants may represent Nth plant designs is uncertain.

Even though the Texaco gasification process is advanced, its costs will be presented separately from the other advanced gasification costs. Based on the above discussion the DuPont and C.F. Braun studies will be used to represent the range of product costs for methanol from Texaco gasification technology, and the Parsons study for Kopper-Totzek technology. The remainder of the studies represent methanol costs for the other advanced gasification technologies. These costs are as follows:

	\$/mBtu	
	Capital Charge Rate	
	11.5%	30%
Koppers-Totzek	7.23	12.42
Texaco	5.90-6.48	9.44-10.41
Advanced Technology	5.30-6.08	8.74-9.78

B. Methanol from Subbituminous Coal

There are two original studies available which investigated the technical and economic feasibility of producing methanol from subbituminous coals. These studies are:

1. "Methanol from Coal: An Adaptation from the Past," Bailey, Davy McKee Corp., 1979.[6]
2. "Research Guidance Studies to Assess Gasoline from Coal by Methanol-to-Gasoline and Sasol-Type Fischer-Tropsch Technologies, Schreiner, Mobil Research and Development Corp., August, 1978.[7]

The Davy McKee study investigated the use of a Davy McKee fluidized bed gasifier, which is a modified Winkler gasifier which has not been demonstrated on a commercial scale. ICI technology was used for methanol synthesis. The Mobil study utilized BGC Lurgi technology for gasification and Lurgi methanol synthesis.

Depth of Design: Neither study seems to have been based on a high level of engineering design. However, since the Davy McKee study utilized modified Winkler/ICI technology and since Davy McKee has designed and built commercial processes using Winkler and ICI technology, their study is probably based on processing and cost correlations associated with plants they have constructed. For the Mobil study, process information was based on either published or licensor data, while investment estimates were principally derived from in-house data. For offsite units vendor quotes were used where obtainable.

Ultimate Analyses of Subbituminous Coal Feedstocks: Ultimate analyses for the subbituminous coals are presented in Table 14. The higher and lower heating values are also shown. Both coals are from Wyoming and have very similar compositions.

Material Balance and Efficiencies: Feedstock and product rates for both studies are presented in Table 15. Methanol and coal are presented on both a short ton per calendar day (tpd) and an energy basis. The Davy McKee study produces 100 percent methanol while the Mobil study produces about 48 percent methanol, 50 percent SNG, and 2 percent naptha. Sulfur, ammonia and coal fines are produced as by-products from the Mobil study. Coal fines are also produced since the Lurgi gasifier cannot process them. By-products were not reported for the Davy McKee study, but are produced; therefore, for economic purposes the sulfur and ammonia yields from the other study were assumed for it.

Product qualities for the Davy McKee study are not reported, but product qualities for the Mobil case are presented in Table 16. The methanol is 99.66 percent pure, but it is still considered to be of fuel grade quality. The SNG is about 96 percent

Table 14

Ultimate Analysis of Subbituminous Coal Feedstocks

<u>Study:</u>	<u>Davy McKee[6]</u>	<u>Mobil[7]</u>
<u>Coal Type:</u>	Wyoming	Wyoming
HHV, Dry, Btu/lb	11,818	11,818
LHV, Dry, Btu/lb	10,963	10,963

Ultimate Analysis of Dry Coal, Wt %

C	69.2	70.8
H	4.7	4.9
O	17.9	18.3
N	0.7	0.7
S	0.4	0.4
Ash	<u>7.1</u>	<u>5.1</u>
Total	100	100
Wt % Moisture (as recieved)	28	28

Table 15

Methanol from Subbituminous Coal:
Feedstock and Product Rates
(50,000 FOEB/CD of Product)

	<u>Davy McKee[6]</u>	<u>Mobil[7]</u>
<u>Mass Basis</u>		
Feedstock		
Dry Coal, tpd	26,820	19,063
Product		
Methanol, tpd	15,227	7,270
Synthetic Natural		
Gas, mscf/CD	-	150
Naptha, bbl/CD	-	1,351
By-products, tpd		
Sulfur	-	63
Ammonia	-	103
Coal Fines	-	1,501
<u>Energy Basis, mBtu/CD, (HHV)</u>		
Feedstocks		
Coal	639,918	450,563
Electricity	3,448	1,198
Products		
Fuel Grade Methanol	295,000	141,388
Synthetic Natural Gas	-	146,588
Naptha	-	7,024
Coal Fines	-	-
<u>Thermal Efficiency, %</u>	45.9	65.3

Table 16

Methanol from Subbituminous Coal: Product Qualities

Davy McKee[6]

The quality of the fuel grade methanol was not reported in the Davy McKee study.

Mobil[7]

1. SNG

<u>Composition</u>	<u>Weight %</u>
Hydrogen	1.7
Methane	95.9
Carbon Dioxide	0.5
Inerts (N ₂ and Ar)	<u>1.9</u>
	100.00

HHV: 975 Btu/scf

LHV: 878 Btu/scf

2. Methanol

	<u>Weight %</u>
CH ₃ OH	99.66
Light Boiling Compounds	0.12
Heavy Boiling Compounds	0.07
Water	0.15

3. Naptha

	<u>Weight %</u>
Gravity, °API	43.5
(R+M)/2 (unleaded)	88.8
Reid Vapor Pressure, lb.	3.5

methane. The naptha product has an octane $((R+M)/2)$ of 88.8, and is a suitable gasoline blending stock.

The thermal efficiencies (based on higher heating values) for the Davy McKee and Mobil cases are 45.9 and 65.3 percent, respectively. This is a very significant difference. The Davy McKee efficiency is a bit lower than the lowest efficiency reported for methanol from bituminous coals (49.3 percent) in Table 9. However, the efficiency for the Mobil case is significantly greater than any of the efficiencies reported for methanol from bituminous coals. One reason for this high efficiency is that the raw syngas from the Lurgi gasifier is high in methane content and the simple isolation of this as product is more efficient than converting it to carbon monoxide and hydrogen and then to methanol. Less processing of the raw syngas is required, and, therefore, a greater efficiency is the result.

Economics: Both studies have been placed on a consistent economic basis as discussed in a previous report.[10a] Table 17 presents the investment costs broken down as much as possible into individual process unit costs. An inspection of Table 17 shows that the total instantaneous investments are \$1.84 billion for the Davy McKee case and \$2.26 billion for the Mobil case.

Operating costs are presented in Table 18. The difference in operating cost between both cases is mainly due to annual coal feedstock cost differences which primarily is a function of process efficiency. As noted earlier, by-product credit for Case 1 is based on the ammonia and sulfur yields of Case 2.

Table 19 and 20 present economic summaries and product costs when using capital charge rates of 11.5 and 30 percent. The methanol product cost for the Davy McKee case ranges from \$6.16-10.26/mBtu while the average product costs for the Mobil case range from \$6.34-\$11.24, depending on the capital charge rate. In addition to average product costs, Tables 19 and 20 present product costs for the methanol, SNG, and gasoline produced in Mobil study which are based on the product value technique discussed in another report.[10a]

It is possible that the capital cost from the Mobil study is more accurate than that from the Davy McKee study; the reason for this is that the original Davy McKee Plant has to be scaled up significantly whereas the other was much closer to the selected 50,000 FOEB/CD. Therefore, the Mobil study's costs will be used in preference.

C. Methanol from Lignite

The following two original studies investigated the technical feasibility of producing methanol from lignite:

Table 17

Methanol from Subbituminous Coals: Capital Cost Summary
Millions of First Quarter 1981 Dollars

	<u>Davy McKee[6]</u>	<u>Mobil[7]</u>
<u>Technology</u>		
Gasification/Methanol Synthesis	Modified Winkler/ICI	Lurgi/ Lurgi
<u>Investment Costs</u>		
Coal Preparation and Handling	87	-
Gasification and Gas Cleaning	98	-
Shift Conversion	36	-
Acid Gas, Sulfur Recovery, Sulfur Guard	175	-
Syngas Compression	66	-
Total: Coal Preparation, Gasification, Processing	462	628
SNG Production	N/A	38
Methanol Synthesis and Distillation	153	102
Oxygen Production	262	161
Offsites and Product Storage	302	451
Infrastructure	17	67
Engineering and Design	119	184
Environmental Studies, etc.	-	3
Other Project Costs	284	331
Contingency	240	295
<u>Total Instantaneous Plant Investment</u>	1840	2260

Table 18

Methanol from Subbituminous Coal: Operating Cost Summary
(Millions of First Quarter 1981 Dollars Per Year)

	<u>Davy McKee[6]</u>	<u>Mobil[7] Case 2</u>
<u>Technology</u>		
Gasification/Methanol Synthesis	Modified Winkler	Lurgi/Lurgi
<u>Raw Materials</u>		
Coal	231	173
Catalysts and Chemicals	8.4	6.9
<u>Utilities</u>		
Power	4.7	2.1
Water	-	-
<u>Labor and Related</u>		
Labor	32.5	49.0
Supplies	33.3	29.0
<u>Capital Related</u>		
Administration and General Overhead	39.4	31.4
Local Taxes and Insurance	59.2	62.6
Interest on Working Capital	7.9	9.8
<u>Gross Annual Operating Cost</u>	416.4	366
By-product Credit	(9.3)	(18.3)
<u>Net Annual Operating Costs</u>	407	348

Table 19

Economic Summary of Methanol from
Subbituminous Coal, CCR = 11.5 Percent
(Millions of First Quarter 1981 Dollars)

	<u>Davy McKee</u>	<u>Mobil</u>
Total Instantaneous Plant Investment	1,840	2,260
Total Adjusted Capital Investment	2,087	2,563
Start-up Cost*	131	163
Pre-paid Royalties	10**	25
Total Capital Investment	2,228	2,751
Initial Catalyst and Chemicals and Working Capital***	131	163
Total Capital Requirement	2,359	2,914
Annual Capital Charge	256	335
Annual Operating Costs	407	348
Total Annual Charge	663	683
<u>Product Cost</u>		
\$/FOEB of Product****	36.33	37.43
\$/mBtu of Product	6.16	6.34
Methanol, \$/mBtu	6.16	7.04
SNG, \$/mBtu	-	5.63
Gasoline, \$/mBtu	-	7.04

* Start-up cost = 6.3 percent of Total Adjusted Capital Investment.

** Royalties were assumed equal to \$10 million unless reported by study.

*** Working Capital and Initial Catalyst and Chemical = 6.3 percent of Total Adjusted Capital Investment.

**** One FOEB = 5.9 mBtu.

Table 20

Economic Summary of Methanol from
Subbituminous Coal, CCR = 30 Percent
(Millions of First Quarter 1981 Dollars)

	<u>Davy McKee</u>	<u>Mobil</u>
Total Instantaneous Plant Investment	1,840	2,260
Total Adjusted Capital Investment	2,053	2,522
Start-up Cost	131	163
Pre-paid Royalties	10	25
Total Capital Investment	2,194	2,710
Working Capital	131	163
Total Capital Requirement	2,325	2,873
Annual Capital Charge	698	862
Annual Operating Costs	407	348
Total Annual Charge	1,105	1,210
<u>Product Cost</u>		
\$/FOEB of Product*	60.55	66.30
\$/mBtu of Product	10.26	11.24
Methanol, \$/mBtu	10.26	12.48
SNG, \$/mBtu	-	9.98
Gasoline, \$/mBtu	-	12.48

* One FOEB = 5.9 mBtu.

1. Production of Methanol from Lignite, prepared by Wentworth Brothers Incorporated (WBI), and C.F. Braun and Company for EPRI.[9]

2. Lignite-to-Methanol, an Engineering Evaluation of Winkler Gasification and ICI Methanol Synthesis Route, prepared by Davy McKee International, Inc.[8]

Both these studies represent approximately the same amount of engineering design. The WBI/C.F. Braun study uses Texaco gasification and WBI methanol synthesis technology. Three cases from this study are presented. Case 1 was prepared by WBI and was designed based on a 55 percent lignite/45 percent water slurry concentration. Gasification of a lignite concentration this high has not been commercially demonstrated. Case 2 represents a C.F. Braun modification of the WBI design still using the 55 percent slurry concentration. Since the 55 percent lignite slurry concentration has not been commercially demonstrated, C.F. Braun also analyzed a methanol from lignite case based on a 43 percent lignite slurry concentration which has been successfully gasified (Case 3).

The DMI study is based on Winkler gasification and ICI methanol synthesis. Both of these technologies have been commercially proven.

Ultimate Analysis of Lignite: All four cases were based on the same lignite, and the ultimate analysis for this lignite is presented in Table 21. Gasifier yields and oxygen requirements for all four cases are based on this analysis.

Material Balance and Efficiencies: Feedstock and product rates for each case are presented in Table 22. Methanol and lignite are presented on both a short ton per calendar day (tpd) and an energy basis. The methanol produced is of fuel grade quality, even though in Cases 1, 2, and 3 the methanol product rates are reported on a dry equivalent basis. All rates are based on 50,000 FOEB/CD of liquid products. Sulfur is the only by-product reported in Table 22.

Thermal efficiencies vary from 43.9 percent for Case 3 to 51.2 percent for Case 1. The 43.9 percent efficiency results from the low lignite concentration in the slurry. The vaporization of the additional water in lower lignite concentration slurries consumes energy in the gasifier and produces larger quantities of synthesis gas. The result is a lower thermal efficiency and an increase of the capacities of all process units except methanol synthesis.

Economics: Table 23 presents all of the investment costs broken down into individual process units. These costs are based on a plant size of 50,000 FOEB/CD of product. An inspection of

Table 21

Ultimate Analysis of Lignite Feedstock

Heating Values

HHV, dry, Btu/lb	10,179
HHV, wet, Btu/lb	9,765
LHV, approximated, dry, Btu/lb	6,460

Ultimate Analysis, Dry Coal, Wt%

C	58.98
H	4.55
O	19.05
N	0.77
S	1.40
Ash	<u>15.25</u>

Total	100.00
Wt.% Moisture (as received)	35

Table 22

Methanol from Lignite: Feedstock and Product Rates
(Normalized to 50,000 FOEB/CD of Product)

<u>Study:</u>	<u>WBI[9]</u>	<u>C.F. Braun[9]</u>		
	<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	<u>Davy McKee[8]</u>
<u>Mass Basis</u>				
<u>Feedstocks</u>				
Lignite, tpd (wet)	44,250	44,596	52,071	48,171
<u>Products</u>				
Methanol (tpd)	15,063*	15,063*	15,063*	15,226
<u>By-Products</u>				
Electricity, energy equivalent per day	8,756	10,107	13,721	-
Sulfur, tpd	324	324	384	312
<u>Energy Basis, mBtu/CD, (HHV)</u>				
<u>Feedstocks</u>				
Lignite	571,705	576,172	671,982	622,363
<u>Products</u>				
Methanol	295,000	295,000	295,000	295,000
<u>Thermal Efficiency, %</u>	51.6	51.2	43.9	47.4

* Methanol on a dry equivalent basis.

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Table 23

Methanol from Lignite: Capital Cost Summary
(Millions of First Quarter 1981 Dollars)

<u>Study:</u>	<u>WBI[9]</u> <u>Case 1</u>	<u>C.F. Braun[9]</u> <u>Case 2</u>	<u>Case 3</u>	<u>Davy McKee[8]</u>
<u>Technology</u>				
Gasification/Methanol Synthesis	Texaco/ WBI	Texaco/ WBI	Texaco/ WBI	Winkler/ ICI
Slurry Concentration, %	55	55	43	N/A
<u>Plant Investment Costs</u>				
Lignite Storage and Preparation	41.1	43.5	49.3	114.2
Syngas Generation, Gas Adjustment, and Puri- fication	637.8	779.4	932.8	--
Gasification, Com- pression and Shift Conversion	--	--	--	278
Acid Gas Removal, Chlorideand Sulfur Guard, Compression	--	--	--	368.7
Methanol Synthesis, Distillation and Hydrogen Recovery	--	--	--	--
Methanol Synthesis	155.7	164.7	164.7	--
Gas Desulfurization, and Sulfur Recovery	23.0	24.4	27.5	35
Air Separation	457.2	483.6	677	--
Utility System	285.96	314.0	314	--
Utilities and Offsites	--	--	--	661.1
General Facilities	154.6	96.0	122.9	
Engineering Fees, Home Office Cost, and License Fees	91.4	92	92.5	225.8
Contingency	263	286	331	219
<u>Total Instantaneous Plant Investment</u>	2110	2283	2628	1901

Table 23 shows that the plant investment estimates vary from \$1.90-2.63 billion. Cases 1, 2, and 3 are based on the same technology. Case 1 was prepared by WBI; whereas Case 2 is C.F. Braun's analysis of the WBI design. Both are based on the same lignite slurry concentration (55 percent).

Braun evaluated the capital costs for adjustments they thought necessary to appraise the WBI work. The necessary adjustments were the addition of one spare gasifier/exchanger set per train and operation with the start-up boiler continuously on the line thus increasing export power. This equipment was added as insurance to maintain production levels and to provide flexibility to the complex. Thus, Case 2 is more conservative; its capital cost is \$170 million more than that for Case 1. It must be noted that to obtain a 55 percent slurry concentration, feed pretreatment is necessary and the cost of pretreatment equipment was not included in either of the estimates.

The instantaneous investment for the C.F. Braun case which utilizes the 43 percent lignite slurry concentration is \$2.63 billion, which is \$350-500 million more than the Case 1 and 2 investments, respectively. This higher investment results from increased capacities of all process units (except methanol synthesis) needed to accommodate the larger amounts of water (and steam) present with the 43 percent lignite slurry.

The total instantaneous investment for the Davy McKee case is \$1.901 billion. This case is based on Winkler/ICI technology.

Operating costs are presented in Table 24. Net annual operating costs range from \$237 million for Case 1 to \$380 million for Case 3. Reasons for the low operating cost estimates for Case 1 are that: 1) general and administration cost have not been included, 2) 1 percent of the total instantaneous plant investment was used for property taxes and insurance as opposed to 2.5 percent for the other studies, and 3) labor costs were reported to be less than those for the other studies.

Operating costs for Case 3 are expected to be higher than those for Case 1 and 2 because of its higher capital investment and higher feedrate of lignite.

Cases 1 and 2 are based on the same technology and lignite slurry concentration. Since Case 2 is a further analysis of Case 1 and is more conservative, it is expected that the operating and capital cost for Case 2 are more representative. Therefore, Case 2 will be used in preference to Case 1 for developing methanol product costs.

Tables 25 and 26 present economic summaries of methanol costs for capital charge rates of 11.5 and 30 percent. For the lower capital charge rate, product costs vary from \$5.70 to \$6.92/mBtu.

Table 24

Methanol from Lignite: Operating Cost Summary
(Millions of First Quarter 1981 Dollars Per Year)

	<u>WBI[9]</u> <u>Case 1</u>	<u>C.F. Braun[9]</u> <u>Case 2</u>	<u>Case 3</u>	<u>Davy McKee[10]</u>
<u>Technology</u>				
Gasification/Methanol Synthesis	Texaco/ WBI	Texaco/ WBI	Texaco/ WBI	Winkler/ ICI
Slurry Concentration, %	55	55	43	N/A
<u>Annual Operating Costs</u>				
<u>Raw Materials</u>				
Coal	160.7	160.7	190	175.7
Fuel	--	12.2	--	--
Catalysts and Chemicals	3.6	3.6	3.6	17.9
<u>Utilities</u>				
Water	--	--	--	0.2
<u>Labor and Related</u>				
Operating	24.5	--	11.2	19.9
Maintenance	--	37.9	45.4	18.4
Administration and Support	--	14.7	16.9	11.5
<u>Capital Related</u>				
Ash Disposal	2.1	2.1	2.5	2.4
Maintenance Materials	52.8	46.3	58.1	27.9
General and Administration	--	24.9	30	25.1
Property Taxes and Insurance	21.1	52.8	65.7	47.5
Interest on Working Capital	8.9	9.6	11.0	8.0
<u>Gross Annual Operating Cost</u>	273.7	376	434.4	354.5
<u>By-product Credit</u>				
Sulfur	(5.8)	(5.8)	(5.9)	(5.7)
Electric Power Export, (3.5¢/kw-hr)	(30.5)	(35.2)	(47.8)	--
<u>Net Annual Operating Cost</u>	237	335	380	349

Table 25

Economic Summary of Methanol from
Lignite, CCR = 11.5 Percent
(Millions of First Quarter 1981 Dollars)

	WBI Case 1	C.F. Braun Case 2 Case 3		Davy McKee
Total Instantaneous Plant Investment	2,110	2,283	2,628	1,901
Total Adjusted Capital Investment	2,393	2,589	2,980	2,156
Start-up Cost	148	160	184	133
Pre-paid Royalties	10	10	10	10
Total Capital Investment	2,551	2,759	3,174	2,299
Working Capital	148	160	184	133
Total Capital Requirement	2,699	2,919	3,358	2,432
Annual Capital Charge	293	317	365	264
Total Annual Charge	530.5	652	745	613
<u>Product Cost</u>				
\$/FOEB of Methanol*	29.07	35.73	40.83	33.61
\$/mBtu of Methanol	4.93	6.06	6.92	5.70

* One FOEB = 5.9 mBtu.

Table 26

Economic Summary of Methanol
from Lignite, CCR = 30%
(Millions of First Quarter 1981 Dollars)

	WBI Case 1	C.F. Braun Case 2 Case 3		Davy McKee
Total Instantaneous Plant Investment	2,110	2,283	2,628	1,901
Total Adjusted Capital Investment	2,355	2,548	2,933	2,122
Start-up Cost	148	160	184	133
Pre-paid Royalties	10	10	10	10
Total Capital Investment	2,513	2,718	3,127	2,265
Working Capital	148	160	184	133
Total Capital Requirement	2,661	2,878	3,311	2,398
Annual Capital Charge	754	815	938	680
Annual Operating Costs	237	335	380	349
Total Annual Charge	992	1,150	1,318	1,029
<u>Product Cost</u>				
\$/FOEB of Methanol*	54.36	63.01	72.23	56.38
\$/mBtu of Methanol	9.21	10.68	12.24	9.56

* One FOEB = 5.9 mBtu.

For the higher capital charge rate, methanol costs vary from \$9.56 to \$12.24/mBtu.

Since the present state-of-the-art gasification of lignite using the Texaco gasifier is costly due to the effect of the lignite water slurry concentration, the Davy McKee costs will be used in preference. The Davy McKee study utilizes the proven Winkler gasification which does not require a coal slurry feed. Thus, the product cost of methanol varies from \$5.70/mBtu for the low CCR to \$9.56/mBtu for the high CCR.

D. Production of Gasoline from Coal via Fischer-Tropsch and Mobil's MTG Technology

There are three original studies available which investigate the technical feasibility of producing gasoline from coal-derived methanol. These studies are:

1. "Coal-to-Methanol-to-Gasoline Commercial Plant," Badger Plants, Incorporated, Cambridge Massachusetts, FE-2416-43-V1,2, March 1979.[10]

2. "Research Guidance Studies to Assess Gasoline from Coal by Methanol-to-Gasoline and Sasol-Type Fischer-Tropsch Technologies," Max Schreimer, Mobil Research and Development Corporation, FE-2447-13, August 1978.[7]

3. "Screening Evaluation: Synthetic Liquid Fuels Manufacture," Prepared by the Ralph Parsons Co. for EPRI.[1]

The Badger study is based on a "slag bath" gasifier which is a new concept and may still require developmental work. (See a more detailed discussion above in Section IV.) Lurgi technology is used for methanol synthesis, and Mobil fixed bed technology is used for methanol-to-gasoline conversion.

The Mobil study actually includes three cases, designated Cases 1, 2, and 3. Cases 1 and 2 utilize Lurgi technology for coal gasification and methanol synthesis, and Mobil fixed bed technology for methanol-to-gasoline conversion. These cases differ in that Case 1 produces approximately 50 percent gasoline and SNG, whereas Case 2 produces approximately 100 percent gasoline. Case 3 uses Lurgi gasification technology's but employs Fischer-Tropsch technology for product synthesis.

The Parsons study is based on the BGC/Lurgi gasifier which still needs to be commercially demonstrated. Fischer-Tropsch technology is used for product synthesis.

Depth of Design: Both the Badger and the Mobil studies are based on a comparable level of engineering design. The investment estimates are of budget or scoping quality. The Badger study is

based on process licensors economic data for proprietary processes and on vendor quotes derived from in-house equipment specifications for non-proprietary processes. Badger states their cost estimate represents an accuracy of minus 5 percent, plus 20 percent. The Mobil study is of the same order of accuracy as the Badger study. The Parsons study is a screening evaluation, and could be expected to be less accurate than the other two studies.

Ultimate Analysis of Coal-to-Gasoline Feedstock: Ultimate analysis for the coal feedstocks are presented in Table 27. The Badger study uses a Southern Appalachian bituminous coal; whereas Mobil uses a Wyoming subbituminous coal, and Parsons uses an Illinois No. 6 bituminous coal. The coal considered in the Badger study is a low sulfur coal which would meet the sulfur dioxide emissions standard for large power plants.[12] It is unlikely that a coal of this quality would be used for synfuels production.

For the Badger study, the coal, free of debris, cleaned, sized, and washed is delivered to the plant site. Thus, this case excludes coal preparation costs which has been included in the other studies.

Material Balance and Efficiencies: Feedstock and product rates for each case are presented in Table 28. All rates are based on 50,000 FOEB/CD of products (excluding by-products). For the Badger study gasoline represents almost 100 percent of the product slate. The efficiency for this case is 49 percent. For Case 1 of the Mobil study the major products are SNG and gasoline and the overall process efficiency is 63.2 percent. The production of SNG increases the overall process efficiency over the all-gasoline cases since the isolation of SNG produced in the Lurgi gasifier requires less energy than gasoline-production. Gasoline is the main product from Case 2; the efficiency of this case is 46.6 percent which is comparable to the Badger plant efficiency. A variety of products are produced from Case 3 (Fischer-Tropsch) with the main products being SNG and gasoline. The efficiency of this case is 57 percent. The efficiency of the Parsons case is 56 percent. While both the Parsons and Mobil (Case 3) studies are based on Fischer-Tropsch technology, their product slates vary widely. For the Parsons case 15,000 FOEB/CD of heavy fuel oil is produced compared to 700 for the Mobil case; whereas 190 mscf/CD of SNG is produced for the Mobil case and only 112 mscf/CD for the Parsons case.

Both of the Fischer-Tropsch synthesis cases produce significant quantities of SNG and/or residual oil. For a transportation fuels oriented synthetic fuels industry, their product slates would be unacceptable. Both cases produce approximately 33 percent transportation fuels (gasoline and diesel fuel). However, currently transportation fuels (jet fuel, diesel fuel, and gasoline) account for 51 percent of the refined petroleum products and this percentage is expected to increase to nearly 55 percent by

Table 27

Coal to Methanol to Gasoline: Ultimate Analysis of Coals

<u>Study:</u>	<u>Badger[10]</u>	<u>Mobil[7]</u>	<u>Parsons[1]</u>
<u>Coal Type:</u>	Southern Appalachian	Wyoming Sub-Bituminous	
HHV, Dry, Btu/lb	12,840	11,818	12,771
LHV, Dry, Btu/lb	-	10,963	11,709
<u>Ultimate Analysis of</u>			
<u>Dry Coal, Wt. Percent</u>			
C	73.8	70.84	69.5
H	4.8	4.85	5.3
O	6.4	18.32	10.0
N	1.6	0.71	1.3
S	1.1	0.43	3.9
Ash	12.3	4.85	10.0
Total	100.0	100.0	100.0
% Moisture (as received)	2.4	28.0	4.2

Table 28

Coal-to-Gasoline: Feedstock and Product Rates
(Normalized to 50,000 FOEB/CD of Product)

	Badger[12]	Case 1	Mobil[7] Case 2	Case 3	Parsons[1]
<u>Feedstocks</u>					
Coal (tpd)	23,147	29,467	37,219	30,406	21,528
Electricity (mBtu/day)	7,572	-	-	-	-
Methanol	-	-	-	4.4	-
<u>Products</u>					
Propane bpd	2,911	1,678	3,606	1,211	-
Isobutane, bpd	4,544	-	-	-	-
Butane, bpd	-	2,380	5,162	160	2,005*
SNG, mSCF/d	-	160	-	190	112
Alcohols, tpd	-	-	-	255	951*
Gasoline, bpd	46,729	23,795	50,843	14,853	10,160
Diesel Fuel, bpd	-	-	-	2,523	6,014*
Heavy Fuel Oil, bpd	-	-	-	680	14,849
<u>By-Products</u>					
Power, mBtu/d	-	470	153	296	-
Coal Fines, mBtu/d	-	22,043	-	-	-
Sulfur, tpd	207	66	83	67	762
Ammonia, tpd	-	111	140	113	-
<u>Energy Basis mBtu/d</u>					
<u>Feedstocks</u>					
Coal	593,497	501,471	633,392	517,454	526,790
Electricity	7,572	-	-	-	-
Methanol	-	-	-	85	-
<u>Products</u>					
LPG	10,897	6,405	13,814	4,619	-
Butane	-	9,975	21,633	690	11,829
SNG	-	157,139	-	190,129	94,524
Alcohols	-	-	-	7,603	5,613
Gasoline	264,855	121,474	259,555	74,607	59,942
Diesel Fuel	-	-	-	13,487	35,484
Heavy Fuel Oil	-	-	-	3,865	87,608
Total	294,552	295,000	295,000	295,360	295,000
<u>Thermal Efficiency, %</u>	49	63.2	46.6	57.1	56.0

* = FOEB

the year 2000 (see Table 7[36]). At the expense of an increased product cost, both of these plants could be altered to meet a more desirable product slate.

Product Qualities: The main products from the Badger and Mobil studies are SNG and gasoline. Chemical and physical analyses of these products are presented in Table 29. Analyses of the products from the Parsons study were not available. The SNG from Cases 1 and 3 of the Mobil study is of satisfactory quality and is compatible with natural gas. The unleaded gasolines presented in Table 29 meet all 1976 ASTM specifications. Compared to typical present-day gasolines these are slightly lower in API° gravity. It would be preferable if the durene content of the gasoline were less than 4 wt. percent since durene contents of 5 wt. percent in conventional gasolines have caused carburetor icing and stalling. The olefinic concentration of the Case 3 gasoline (20 vol. %) is higher than that of conventional gasolines, and experience with higher olefinic gasolines is limited. Consequently, marketing such a gasoline would require further testing.

All of the propane and butane products are satisfactory fuels. The isobutane from the Badger study is of high purity and may be used as a petrochemical or as a refinery feedstock. The diesel fuel from Case 3 of the Mobil study could be marketed as a premium diesel fuel, No. 1-D. The heavy fuel oil from this case contains no sulfur or metals and thus could be marketed as a premium gas turbine fuel. The alcohols from these case are a mixture of C₂-C₆ alcohols, and are essentially free of acids, aldehydes, ketones and water.

MTG Process Economics: Table 30 presents capital investment costs broken down into individual process unit costs. An inspection of this table shows that the estimates of the total instantaneous plant investment for the MTG process range from about \$2.6 billion for the Badger study and Case 1 of the Mobil study, to about \$3.6 billion for Case 2 of the Mobil study. The Badger study and Case 2 of the Mobil study are both designed to produce gasoline as the major product. Since both of these cases are based on Mobil's methanol-to-gasoline technology and produce similar product slates, it is expected that their investment costs would be comparable, but this is not the case. Mobil's capital estimate is nearly \$1 billion more than Badger's. Even though the capital cost of a subbituminous coal plant is expected to be greater than a bituminous coal plant, this difference is much too large. Table 30 shows that the "gasification, et al" costs for these cases are almost identical, even though the Mobil case is slightly less efficient and operates with subbituminous coal as a feedstock as opposed to bituminous coal for the Badger study. On this basis one would expect Mobil's gasification costs to be greater than Badger's, which would tend to make the investment difference between the two studies even greater.

Table 29

Product Qualities: Coal-to-Gasoline

<u>Study:</u>	<u>Case 1</u>	<u>Mobil[7]</u>		<u>Case 3</u>
1) <u>SNG</u>				
<u>Composition, %</u>				
Hydrogen	1.7			3.8
Methane	95.5			89.7
Ethene	-			1.0
Ethane	0.2			2.3
Propene	-			1.0
Propane	0.1			0.1
Butane	0.1			-
Carbon Dioxide	0.5			0.5
Inerts (N ₂ + Ar)	1.9			1.6
Total	100			100
Heat of Combustion, Btu/scf	980			1000
Water	0.01%			0.01%
Sulfur	None			None
Carbon Monoxide (0.1% Max)	0.02%			0.07%
<u>Study:</u>	<u>Badger</u>	<u>Mobil[7]</u>		<u>Case 3</u>
		<u>Case 1,2</u>		
2) <u>Gasoline</u>				
Gravity, °API	62.7	61.4		67.2
Research Octane Number	92.7	93		91
Motor Octane Number	82.7	83		83
Volatility				
Reid Vapor Pressure, lb.	10.0	10.0		10.0
Distillation, °F				
IBP	79	85		86
10%	106	110		108
30%	140	146		137
50%	187	200		186
70%	259	262		249
90%	339	336		335
EP	390	388		420
Sulfur, Wt.%	Nil	Nil		Nil
Composition, Vol. %				
Paraffins	54	51		60
Olefins	12	11		20
Napthenes	7	9		3
Aromatics	27	29		17
Durene Content	4 Vol.%	4.6 Wt.%		-

Table 30

Coal-to-Gasoline: Capital Cost Summary
(Million of First Quarter 1981 Dollars)

<u>Study:</u>	<u>Mobil[7]</u>				
	<u>Badger[10]</u>	<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	<u>Parsons[1]</u>
<u>Technology (Gasification/ Synthesis)</u>	<u>Slag Bath/ Lurgi/MTG</u>	<u>Lurgi/ Lurgi/MTG</u>	<u>Lurgi/Lurgi/ MTG</u>	<u>Lurgi/ Fischer- Tropsch</u>	<u>BGC Lurgi/ Fischer- Tropsch</u>
<u>Investment Costs</u>					
Coal and Lime Preparation	78	-	-	-	41
Coal Gasification	210	304	-	307	147
Shift Conversion	63	49	-	49	57
Acid Gas Removal	216	109	-	110	244
Sulfur Recovery	18	89	-	90	44
Hydrocarbon Recovery	-	-	-	30	-
By-product Separation and Recovery	-	78	-	19	100
Syngas Compression	36	-	-	-	70
Methanol Synthesis	220	82	-	-	-
Cyrogenic Hydrogen Recovery	18	4	-	9	31
Total Gasification, et al	859	715	840	614	733
SNG Production	-	38	-	37	-
Gasoline Production	216	106	370	-	-
F-T Synthesis and F-T Product Processing	-	-	-	218	342
Oxygen Production	351	167	261	168	266
Steam and Power Generation	121	231	266	274	11
Cooling Water and Make-up, WWT	90	69	128	77	-
Environmental	63	-	-	-	-
Waste Disposal	-	104	-	109	-
Storage and Shipping	11	34	-	13	-
General Facilities	118	-	286	-	-
Infrastructure	-	70	83	71	-
Other Project Costs	-	376	510	270	-
Engineering and Design	246	189	288	208	-
Miscellaneous	138	107	-	169	212
Sub-Total	2197	2206	3022	2228	1624
Contingency	330	331	455	343	244
Contractor's Fee	56	55	76	57	41
Total Instantaneous Plant Investment	2583	2592	3563	2628	1909

Operating costs are presented in Table 31. Operating costs for the Mobil Case 2 study are \$150 million more than those for the Badger study. Unfortunately, not enough information was available to reconcile these capital and operating cost differences.

Tables 32 and 33 present economic summaries and average product costs when using capital charge rates (CCR) of 11.5 and 30 percent. The average product costs for the MTG processes range from \$7.37-9.75/mBtu for the low CCR and from \$12.94-17.43 for the high CCR. In addition to average product costs, product costs for the various studies based on the product value method discussed in a previous report are also presented in these tables.[10a]

While the cost estimates of these two studies are difficult to reconcile, the incremental product cost to produce gasoline from methanol using the MTG process can be determined and may be more consistent. To accomplish this the incremental investment and operating costs will be determined between: 1) the Badger methanol plant (from the bituminous coals section) and the Badger gasoline from methanol plant, and 2) the Mobil methanol plant (from the subbituminous coals section) and the Mobil (Case 1) gasoline from methanol plant. Then the incremental product costs for each case will be compared. Mobil's Case 1 MTG unit is sized to produce 20,600 FOEB/CD of gasoline while Badger's was sized to produce 45,000 FOEB/CD. Therefore, in the economics to be presented below the Mobil study MTG unit has been scaled up to 45,000 FOEB/CD.

The incremental instantaneous plant investments are:

1. \$634 million - Badger Study
2. \$596 million - Mobil Study

The incremental operating costs are:

1. \$97 million - Badger Study
2. \$53 million - Mobil Study

After determining the total annual charge per the procedure discussed in a previous report,[10a] the incremental charge to produce 45,000 FOEB/CD of gasoline (50,000 FOEB/CD of total product) from methanol via the Mobil MTG process is:

	<u>\$/mBtu</u>	
	<u>Capital Charge Rate</u>	
	<u>11.5%</u>	<u>30%</u>
Badger	1.76	3.12
Mobil	1.45	2.87

Table 31

Coal-to-Gasoline: Operating Cost Summary
(Millions of First Quarter 1981 Dollars Per Year)

	<u>Badger[10]</u>	<u>Mobil[7]</u>			<u>Parsons[1]</u>
		<u>Case 1</u>	<u>Case 2</u>	<u>Case 3</u>	
<u>Raw Materials</u>					
Coal	232	184	232	186	216
Limestone	8.4	-	-	-	-
Catalysts and Chemicals	27.3	7.9	-	9.4	11
<u>Utilities</u>					
Power	28.3	-	-	-	-
Water	-	2.2	-	2.2	-
<u>Labor and Related</u>					
Operations	19.5	9.6	-	14.2	-
Maintenance	13.4	46.5	-	53.1	-
Supervision	1.7	-	-	-	-
General Services	3.1	-	-	-	-
<u>Capital Related</u>					
Operating	-	2.4	-	3.5	-
Maintenance	38	31.1	-	35.5	-
Administration and General Overhead	-	34.6	-	41.7	-
Local Taxes and Insurance	7.5	71.9	-	80	-
Interest on Working Capital	5.2	5.3	8.2	5.5	7.0
<u>Other Operating Cost</u>	32	-	295	-	135
<u>Gross Annual Operating Cost</u>	416	396	535	431	
By-product Credit	(3.8)	(17.4)	(11.0)	(9.6)	(13.9)
<u>Net Annual Operating Cost</u>	412	378	524	422	355

Table 32

Coal-to-Gasoline: Economic Summary, CCR = 11.5%
(Millions of First Quarter 1981 Dollars)

	Badger	Mobil			Parsons
		Case 1	Case 2	Case 3	
Total Instantaneous Plant Investment	2583	2592	3563	2688	1904
Total Adjusted Capital Investment	2929	2939	4040	3048	2159
Start-up Cost	182	181	249	188	133
Pre-paid Royalties	26	26	34	20	9
Total Capital Investment	3136	3146	4323	3256	2301
Working Capital	182	181	249	188	133
Total Capital Requirement	3319	3327	4572	3444	2434
Annual Capital Charge	382	383	526	396	280
Annual Operating Costs	412	378	524	422	355
Total Annual Charge	794	761	1050	818	635
<u>Average Product Cost</u>					
\$/FOEB of Product	43.51	41.68	57.52	44.83	34.79
\$/mBtu of Product	7.37	7.06	9.75	7.60	5.90
<u>Product Costs, \$/mBtu</u>					
LPG	5.82	6.17	7.72	6.56	5.31
Butane	5.82	6.17	7.72	6.56	5.31
SNG	-	6.41	-	6.82	5.52
Alcohols	-	-	-	8.52	6.90
Gasoline	7.55	8.01	10.03	8.52	6.90
Diesel Fuel	-	-	-	7.67	6.21
Heavy Fuel Oil	-	-	-	6.56	5.31

Table 33

Coal-to-Gasoline: Economic Summary, CCR = 30%
(Millions of First Quarter 1981 Dollars)

	Badger	Case 1	Mobil Case 2	Case 3	Parsons
Total Instantaneous Plant Investment	2,583	2,592	3,563	2,688	1,904
Total Adjusted Capital Investment	2,883	2,893	3,976	3,000	2,125
Startup Cost	182	181	249	188	133
Pre-paid Royalties	26	26	34	20	20
Total Capital Investment	3,091	3,100	4,259	3,208	2,278
Working Capital	182	181	249	188	133
Total Capital Requirement	3,273	3,281	4,508	3,396	2,411
Annual Capital Charge	982	984	1,352	1,019	723
Annual Operating Costs	412	378	524	422	355
Total Annual Charge	1,394	1,362	1,876	1,441	1,078
<u>Average Product Cost</u>					
\$/FOEB of Product	76.37	74.65	102.82	78.95	59.09
\$/mBtu of Product	12.94	12.65	17.43	13.38	10.01
<u>Product Costs, \$/mBtu</u>					
LPG	10.19	11.06	13.80	11.36	9.04
Butane	10.19	11.06	13.80	11.36	9.04
SNG	-	11.49	-	11.80	9.39
Alcohols	-	-	-	14.75	11.74
Gasoline	13.25	14.36	17.93	14.75	11.74
Diesel Fuel	-	-	-	13.28	10.57
Heavy Fuel Oil	-	-	-	11.36	9.04

The costs from the Badger study are slightly higher than those from Mobil. Since Mobil has researched and developed the MTG process, it is believed that their study is more reliable; therefore, their costs will be used in preference to Badger's.

Fischer-Tropsch Process Economics: This section examines the investment cost differences between the two Fischer-Tropsch studies. The instantaneous plant investment of the Mobil/Fischer-Tropsch case is \$719 million more than that of the Parsons case. When inspecting onsite process equipment costs, the Parsons case cost \$343 million more. However, the cost of offsite type equipment is \$908 million more for the Mobil case. Therefore, even though there is a large onsite investment cost difference, the major differences between the two studies appears to be in offsite investment costs. The Mobil study is probably more accurate since it is based on a more thorough design. Operating costs from the Mobil/Fischer-Tropsch study are \$75 million greater than those from the Parsons study. Unfortunately not enough information was available to reconcile these differences.

Tables 32 and 33 present economic summaries and average product costs for both CCR's. The average product costs for the two Fischer-Tropsch studies ranged from \$5.90-7.60/mBtu for the low CCR to \$10.01 to 13.38/mBtu for the high CCR. Product costs based on the product value method are also presented in these tables. Since the Parsons' study is based on a less thorough design than the Mobil study, the Parsons' study will not be further investigated.

To determine the average product cost difference between a Fischer-Tropsch synthesis plant and a methanol synthesis plant, the Mobil study (Lurgi gasification/Fischer-Tropsch synthesis) can be compared with Mobil's Lurgi gasification/Lurgi methanol synthesis case from Table 17. Differences in investment and operating costs between these two cases reflect the differences in synthesis technology. The instantaneous plant investment difference is \$355 million and the operating cost difference is \$67 million with the Fischer-Tropsch case being greater. These figures translate into an average product cost difference of \$1.00/mBtu.

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